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COLOR REMOVAL AND SLUDGE DISPOSAL PROCESS FOR KRAFT MILL EFFLUENTS

Edgar L. Spruill, Jr.

Continental Can Company, Incorporated

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About 12% BOD_5 reduction was observed, and average TOC reduction was nearly 40%. The chief negative factor is need for emergency protection against alkaline impact on secondary treatment and receiving stream.

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Primary clarification and sludge disposal are included in the process. Operating costs, exclusive of capital factors, are estimated at \$0.50-\$0.80 per ton of paper, or 5.5¢ to 6.5¢ per thousand gallons, depending on fiber losses and water usage. (EPA)

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EPA-660/2-74-008
February 1974

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By

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Program Element 1B2037

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About 12% BOD_5 reduction was observed, and average TOC reduction was nearly 40%. The chief negative factor is need for emergency protection against alkaline impact on secondary treatment and receiving stream.

Following centrifuge dewatering, sludge incineration has had minimal impact on kiln operation; there were some adverse effects on lime quality. Lime recovery was 93%. Mill kiln capacity must be increased about 25%.

Primary clarification and sludge disposal are included in the process. Operating costs, exclusive of capital factors, are estimated at \$0.50-\$0.80 per ton of paper, or 5.5¢ to 6.5¢ per thousand gallons, depending on fiber losses and water usage.

This report was submitted in fulfillment of Project Number 12040 DRY by the Continental Can Company, Inc., Mill Operations Division, Hodge, Louisiana, under the sponsorship of the Environmental Protection Agency.

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SECTION I

CONCLUSIONS

A treatment process, employing lime in concentrations below the solubility limits of calcium hydroxide in water, has been developed for removal of colored matter from unbleached kraft mill effluent and for recovery of the lime and precipitated organic substances. A full-scale plant embodying this process has been designed and built at Continental Can Company's Hodge, Louisiana, mill and operated since August 18, 1971. Operation through August 31, 1972 is described in this report, and the following conclusions are drawn:

1. Effectiveness of the Process

   - Addition of lime to combined, unclarified kraft pulp and paper mill effluent, at rates below 1100 mg/l as CaO, can produce average color reduction of 80% on a continuous, mill-scale basis.

   - If system reliability can insure uninterrupted lime feed at rates which maintain a residual dissolved calcium concentration of at least 400 mg/l, color reduction will exceed 85% and range to above 90%. The required lime addition will normally be about 1000 mg/l as CaO. (High sodium alkalinity may reduce this performance.)

   - A reduction of about 12% in BOD_5 is realized, and removal of solid-phase organic material is virtually 100%. However, BOD_5 may actually increase if lime feed is inadequate or intermittent.

   - Where NSSC materials contribute more than half of the color, efficiency of color removal may be reduced 15% or more.

   - Sedimentation of calcium carbonate after carbonation is adversely affected by incomplete removal of colored substances and is influenced by other factors not fully elucidated in this study.

   - Precipitated color bodies, together with the usual "primary treatment" sludge, can be disposed of by incineration in a lime kiln, coincidentally with recovery of the lime utilized in the treatment.

2. Design and Performance of Equipment

   - Basic concepts and criteria used for design of most equipment for the project have proved valid.

- Kiln gas flow rate provided for carbonation must reflect minimum CO_2 concentration which is likely to occur, and must include the demand imposed by sodium alkalinity of the effluent.

- Rake torque imposed by the carbonation clarifier is below expectations based on other calcium carbonate slurries. A greater area and retention time allowance for this clarifier may be justified.

- Centrifuge dewatering of sludges requires careful choice and application of equipment, but the centrifuge appears to represent the method of choice for this operation.

3. Effects on Kraft Mill Operations

- For kraft mills treating 10,000 - 15,000 gallons of effluent per ton of pulp, lime kiln requirements are increased about 25%.

- Where filler clays and similar insolubles are avoided, lime contamination will remain within tolerable limits.

- Requirements for carbon dioxide supplied by lime kiln stack gas impose a constraint upon the scheduling of interruptions in kiln operation, as do lime supply and sludge disposal.

- Holding ponds or other emergency alternatives are required to prevent alkali damage to secondary treatment operations (or to receiving stream) in case of system failure at the carbonation stage.

4. Operating Costs

- To treat 13 million gallons of effluent per day from a 1,500 tons-per-day, integrated kraft pulp and paper mill, operating costs will be approximately $0.50 per ton of paper (assumed fuel cost, 48¢ per million Btu). Costs of depreciation, insurance and taxes are not included.

- The functions of conventional primary clarification and final sludge disposal are provided within the stated costs of the process.

SECTION II

RECOMMENDATIONS

There is evidence that, after color precipitation with lime, the final effluent may extract color from natural organic materials during and after secondary treatment. This phenomenon was outside the scope of this project, but the mechanisms and magnitude of such color effects on receiving streams should be explored.

The effect of lime kiln stack gas on color, total organic carbon and BOD_5 of effluent needs further study.

The effects of the organic components of color sludge on the heat requirements for calcining have not been adequately measured. Study might properly include direct establishment of accurate heat and material balances, as well as examination of pyrolysis products in the exit gases from the kiln.

It should be ascertained whether moderate further increases in lime dose have significant effects on reduction of TOC and BOD_5.

SECTION III

INTRODUCTION

The Kraft Effluent Color Problem

The spent liquors from wood pulping by alkaline methods, such as the kraft process, are very dark colored; indeed they are commonly referred to as "black liquor." Further delignification procedures involving alkalis, such as caustic extraction steps in bleaching, also yield highly colored solutions. The kraft process characteristically provides for separation, concentration and combustion of spent liquor to recover soda chemicals and heat value; however, existing technology does not prevent some loss, and effluents from kraft mills are distinctly colored.

Discharge of substantial ratios of untreated mill effluent into streams has long been recognized as dangerous to fish and other aquatic life. The adverse effects are largely alleviated by adaptations of the conventional processes of "primary" and "secondary" waste treatment. Although such treatment may be very effective in removal of suspended solids and reduction of BOD, these procedures are almost totally ineffective in reducing color of the effluent.

Color contributed by paper mill effluent is at least esthetically undesirable, and it may make the stream unsuitable for such uses as municipal water supply or certain recreational purposes. Residual organic substances, refractory to bio-oxidation, may represent a significant COD (or TOC) value.

Previous Work

In 1952, an investigation and review revealed that much work had been done to develop means for reducing color of pulping effluents, but that no practical methodology had resulted, and little technical literature on the subject existed. (1) It was found that color could be substantially removed from solution by a number of agents, including alum, mineral acids, ferric sulfate, lime, barium aluminum silicate, activated silica, and various heavy metal salts, and by combinations of these agents. However, the costs of such treatments, applied to the great volume of water discharged by a pulp and paper mill, were very high, and disposition of the residues presented further problems for which there were no attractive solutions.

This review marked the beginning of a strong and sustained effort, supported by the paper industry, to develop technology for elimination or reduction of color from chemical pulping and bleaching effluents. Efforts were largely directed toward use of lime, both because that material is low in cost and because it offers possibilities for recovery and re-use in the pulping operation. The ensuing program produced a number of reports showing that lime could yield a good degree of color reduction and presenting several attempts at practical recovery of the lime. (2,3,4,5,6,7)

The combination of the lime treatment research program of the N.C.S.I. (National Council for Stream Improvement, more recently changed to N.C.A.S.I., adding "Air and . . ."), is a patented process fully reported in their Technical Bulletin No. 157. (8,9) That process was specifically designed for bleaching system effluents (especially effluent from caustic extraction stages). The method involves use of very large amounts of lime (typically about 20,000 parts, w/w, per million parts of effluent), and the colored substances are said to be deposited on suspended, solid-phase calcium hydroxide. The color laden calcium hydroxide is sedimented in a clarifier and withdrawn as a heavy sludge. The treated effluent is carbonated to remove soluble lime, converting it to calcium carbonate, which is then removed and mixed with the calcium hydroxide sludge. The mixed sludge is filtered and is then used to causticize sodium carbonate in the kraft mill liquor-processing system. Color bodies are dissolved into the kraft cooking liquor and ultimately arrive with the spent liquor at the recovery furnace, while the lime is conventionally reburned and re-used.

The N.C.S.I. scheme, called "The Massive Lime Process" for evident reasons, has the advantages of providing a high degree of color removal, a large reduction in COD, a substantial reduction in BOD_5 (at least from bleachery wastes), and a means for disposing of the separated color bodies and recovering the lime. Its chief disadvantages are: the very large amount of lime required and the substantial changes in operation of the kraft causticizing system. Since no more lime can be employed than is needed for causticizing, the volume of water which can be treated is limited to 3,000 to 5,000 gallons per ton of kraft pulp made.

The Massive Lime Process has been studied under EPA grant 12040 DYD, for which a comprehensive report has been prepared. (10)

Considerable work has been done by at least one investigator to develop a color removal process based on commercial alum. (11) Although good color reduction has been reported and a sludge recovery procedure has been outlined, economic feasibility does not appear to have been developed.

The use of activated carbon, especially as a "polishing" treatment to remove the last traces of color (12,13) has been studied, and the possibilities for re-use of the reclaimed and renovated water has been discussed. (14)

Color Precipitation by Below-Saturation Lime Concentrations

Technical personnel at the Hodge, Louisiana mill of Continental Can

Company, Inc. maintained active contact, through corporate membership in N.C.S.I., with the N.C.S.I. color removal research efforts. After several years consideration of the problem of adapting the Massive Lime Process to their need to treat a rather large volume of effluent, a laboratory program was begun exploring lime and other precipitants.

Data developed by Berger and others of N.C.S.I. was subjected by them to analysis which led them to the conclusion that lime precipitation of color, as they had practiced it, proceeds by a mechanism wherein adsorption of color bodies on solid calcium hydroxide is an essential factor. (15,16)

Observations made at Hodge led to the hypothesis that soluble calcium hydroxide could effect color precipitation at satisfactory rates by utilizing other solid-phase surfaces for deposition. It was further reasoned that such a material, or combination of materials, might confer needed thickening and dewatering properties.

The ensuing program involved addition of various solids, co-precipitation of solids, use of flocculation aids, and a variety of mechanical and chemical modifications of treatment.

In the course of the study, it was found that, in the presence of the amounts of fiber "fines" commonly escaping paper mill reclaim systems and passing into the sewer, color could be rapidly precipitated with lime additions well below the normal water solubility of calcium hydroxide. The economically fortuitous availability of this material led to intensive study of the properties of the precipitates as related to the precipitation conditions and the implications of various flowsheet possibilities.

It was observed that samples representative of total kraft mill effluent, when treated with 1,000 to 1,500 mg/l of CaO, exhibited settling rates more rapid than simple primary clarification of the original effluent. The precipitates could be thickened to volumes suitable for further dewatering by vacuum filtration or centrifugation. Direct vacuum filtration proved unpromising. However, when these same precipitates were mixed with recausticizing sludge (calcium carbonate) in proportions approximating the normal availability of each, acceptable filtration properties were found. Formation rates from 35 lbs./sq. ft./hr. to several times that value were observed, depending upon the amounts of precipitated lignin and of fiber in the mixture.

Centrifuge dewatering efforts were made, using a laboratory centrifuge with 50ml glass bottles. Because of the elastic nature of the fiber-containing sludge, and because minimum cake volume could not be observed or measured under dynamic conditions, no consistent or conclusive data were obtained.

A small (2 g.p.m.) pilot system, providing continuous lime addition, sedimentation and carbonation, was constructed to test the validity of bench-scale data against the variations of effluent quality, and to compare continuous settling with laboratory batch tests. The results indicated consistent color reduction of kraft effluents in the range of 80 - 90%. The presence of NSSC (hardwood) process effluent reduced this performance, but at reasonably low ratios, the effect was not serious. Sedimentation experience seemed to support previous estimates of permissible rise rates around 1.0 g.p.m./ft.^{2}

Project Definition

Progress of the research work at Hodge had been shared with the Executive Secretary of the Louisiana Stream Control Commission. From an early date, the possibility of a treatment facility to handle total (or nearly so) mill effluent was envisioned, to deal with the special color considerations of Hodge mill discharge into a low-flow stream.

The cost of so expensive a venture with unproven technology was forbidding. Since the most serious problems lay in the realm of integrating the system with mill lime processing, a system large enough to establish practicality -- but too small to serve the mill -- seemed even less attractive. Since a full-scale treatment facility should have wide implications in the advance of industrial waste treatment technology, a conference was arranged between representatives of Continental Can Company and the Federal Water Pollution Control Administration (later, Federal Water Quality Administration, and then incorporated into the Environmental Protection Agency). The possibility of a demonstration grant for technical studies, with some participation in construction costs, under the FWPCA program or technology development, was the subject of this meeting and others to follow.

The ensuing discussions, and the agreement of the Continental Can Company corporate management to undertake the financial commitment, led to the offer and acceptance of a demonstration grant with the following plan and objectives:

Beginning with the tentative flow sheet shown in Figure 1, laboratory and pilot plant work would be extended and refined, and a facility would be designed and built, capable of treating the entire process effluent of the Hodge mill.

Upon completion of construction, the system would be placed in operation, and for a period of one year studies would be carried out and operational effort would be made to:

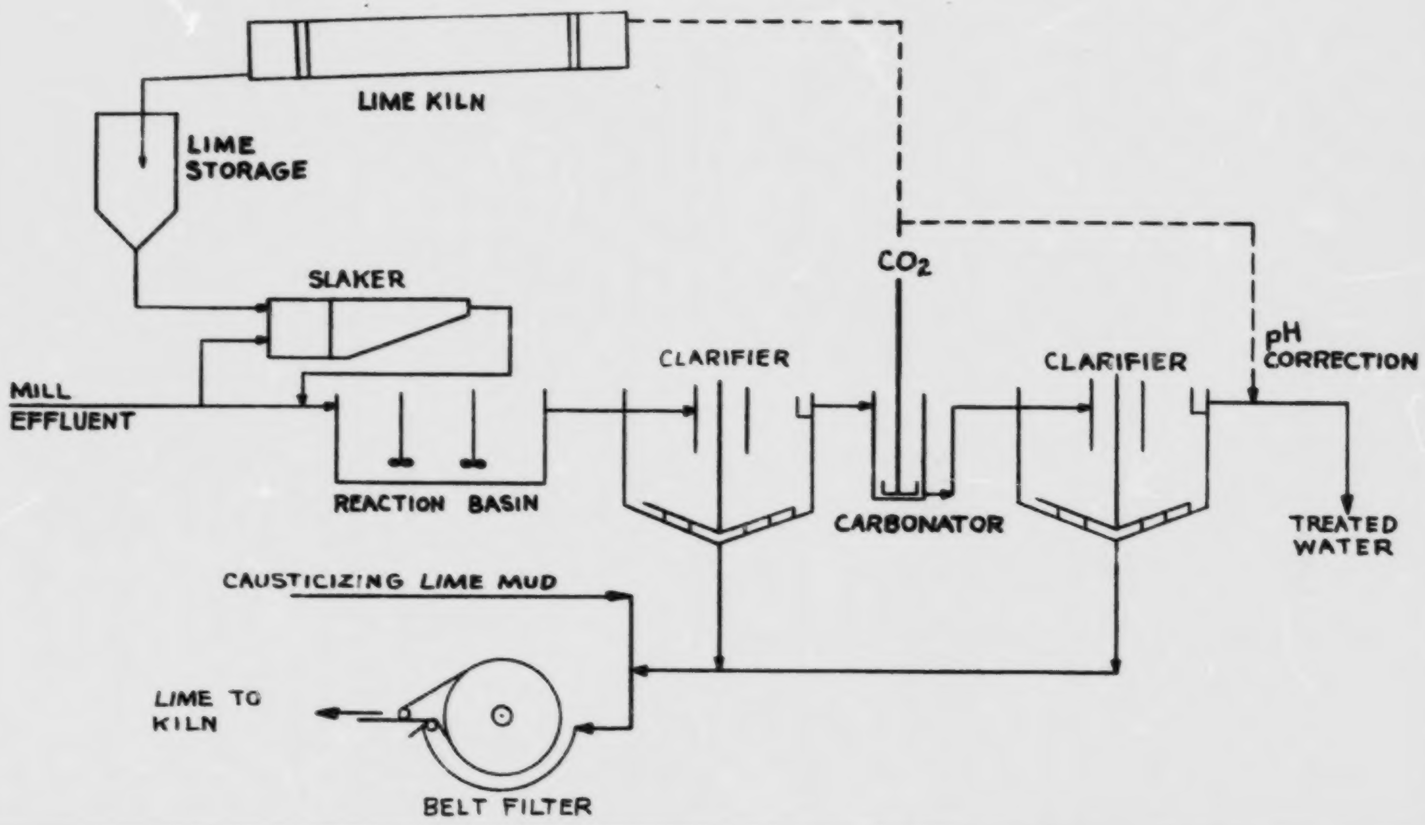
1. Demonstrate effectiveness of the process in reducing color of pulp and paper mill effluents.
2. Demonstrate practicability of the process as a means for disposal of the fibrous sludge normally separated in paper mill primary clarification systems.

3. Determine effectiveness of the process in reducing oxygen demand of pulp and paper mill effluents.

4. Assess the effects on lime calcining of using a lime kiln to incinerate the combined sludges (including fiber) from the lime treatment system.

5. Develop economic parameters related to the process.

It was agreed that a continuous study program would be maintained throughout the demonstration period to record related variations of water papameters (pH, color, BOD, COD or TOC, calcium), sludge properties and lime kiln operations. Interpretations would be developed and the study expanded or modified to gain broader understanding in pertinent areas.


LIME KILN
LIME STORAGE
SLAKER
MILL EFFLUENT
REACTION BASIN
CLARIFIER
CO_2
CARBONATOR
CLARIFIER
pH CORRECTION
TREATED WATER
CAUSTICIZING LIME MUD
LIME TO KILN
BELT FILTER

PROCESS FLOW SHEET - EARLY VERSION

Figure 1

SECTION IV

DEVELOPING A PLANT DESIGN

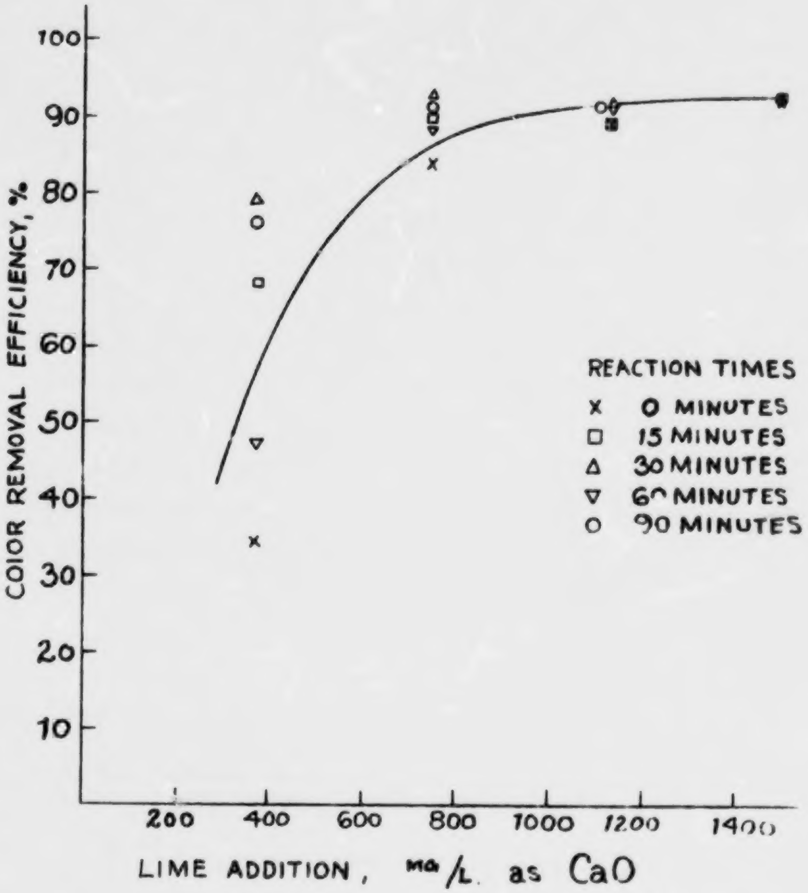
Bench Studies

A renewed program of bench-scale study of the process was instituted to confirm the applicability and effectiveness in treating the range of effluent variations to be expected in plant operation; to define more closely the limits of the process; and to provide data for engineering the system.

Color Removal Efficiency. A series of jar tests was conducted to examine the effect of the lime dosage and reaction time on color removal efficiency. The results are plotted in Figure 2. It appears that at a dosage of 1,000 mg/l $Ca(OH)_2$ and greater, reaction time is not important. The benefit of adding amounts of lime greater than 1,000 mg/l is small, in this case. The original color of this effluent was 300 units, which is fairly low, and the source is a combined screen room and paper machine sewer.

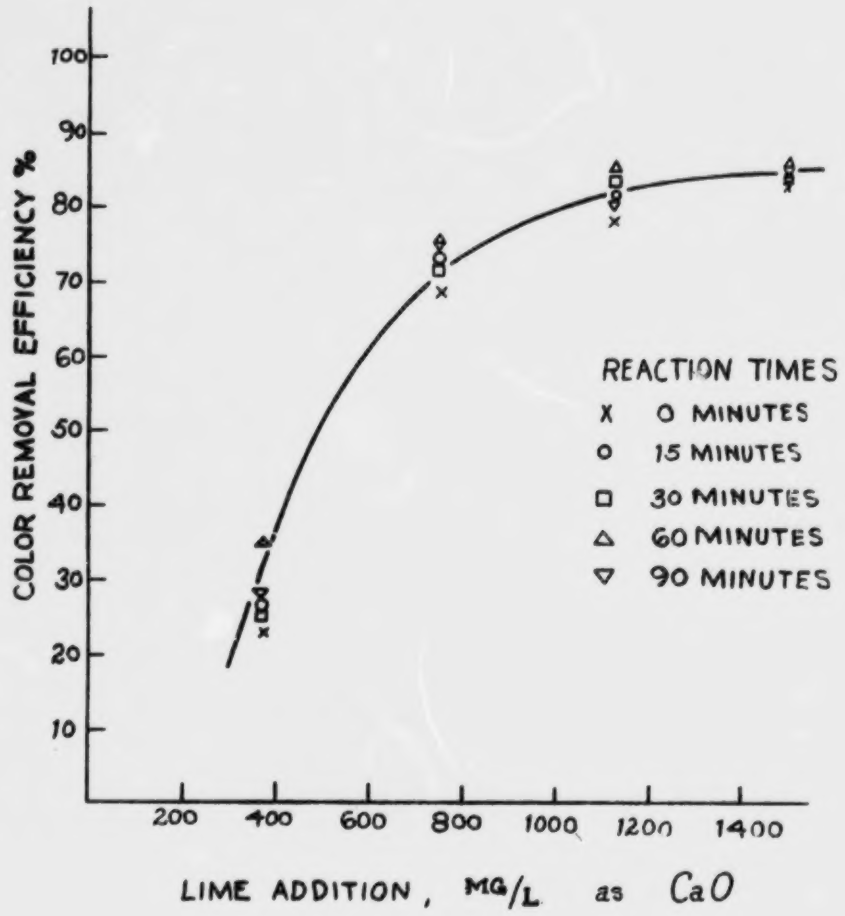
A similar series was conducted with additions of kraft black liquor and NSSC brown liquor to simulate total mill effluent of an integrated mill. The mixture had a color of 550 units, representing the equivalent of a mill with good pulp washing and low recovery losses. Treatment with lime at several concentrations and reaction times was employed to generate the data plotted in Figure 3. It is apparent that at any level of lime dosage, the efficiency of color removal is lower than in the previous series. The effect of reaction time still appears to be small. Using 1,500 mg/l $Ca(OH)_2$ (or 1,135 mg/l CaO), an 85% color reduction was observed. It will be noted that this curve does not appear to extrapolate to zero on the abscissa; this is largely due to carbonates in the effluent, which precipitate some of the calcium ion.

Another series of tests were made using effluent obtained while the mill was producing and using some NSSC as well as kraft pulp. The color of the effluent was 2,820 units, which is in the high range, and the predominant contribution of color was from the semi-chemical process. At a lime (calcium hydroxide) dosage of 1,600 mg/l, a 68% color removal was achieved, and at that dosage increasing the reaction time from 30 minutes to 60 minutes had no effect. These data are shown in Table 1.



EFFECT OF TIME AND LIME DOSAGE ON COLOR (KRAFT)

Figure 2



EFFECT OF TIME AND LIME DOSAGE ON COLOR (WITH NSSC)

Figure 3

TABLE 1

EFFECT OF VARIABLES ON % COLOR REMOVAL EFFICIENCY (WITH NSSC)

Treatment Time, min	Lime Addition as $Ca(OH)_2$ 1,200 mg/l	1,600 mg/l
30	18.4	68.1
60	41.5	68.0

Several trials were made in which black liquor and brown liquor were added to a kraft screen room and paper mill effluent. "Normal" (N) amounts of these liquors were estimated, to represent proportional waste inputs of these materials from all mill sources. Twice-normal amounts of black and brown liquors were added to determine their effect. In these series, the effluent samples were treated with 1,500 mg/l $Ca(OH)_2$ for 30 minutes. The results are presented in Table 2. They indicate that about 90% color removal is achieved on effluent to which no additions were made. Addition of 2N black liquor had little effect on the color removal efficiency. However, the data in sets "B" and "C" suggest that the color of the added brown liquor is considerably more difficult to remove, and that black liquor may actually assist in removal of NSSC color.

TABLE 2

EFFECT OF ADDED BLACK AND BROWN LIQUOR ON % COLOR REMOVAL EFFICIENCY

Sample	Black Liquor	Brown Liquor Added None	2N
A	none	88	85
A	2N	86	82
B	none	92	57
B	2N	90	79
C	none	90	49
C	2N	89	59

All of the work described above was done by "jar-test" techniques. In addition, a number of runs were made with the 2 g.p.m. pilot plant referred to in Section III. One series yielded the data shown in Table 3. The pilot plant was operated with 20 minutes retention in the reactor and rise rates of about 1.0 g.p.m./ft.^{2} in the clarifiers. Retention time in each clarifier was also about 20 minutes. For this series, feed effluent was stored in a tank, so that feed for each run would be constant in composition. The series contains some feed samples darkly colored due to the presence of brown liquor. It appears that color reduction efficiencies in excess of 80% are possible if sufficient lime is added.

TABLE 3

PILOT PLANT COLOR REMOVAL DATA

Input Color	Output Color	% Color Removal	mg/l CaO Added
850	90	90	1,000
900	60	93	1,000
4,000	2,300	42	1,000
4,750	825	83	2,000
1,950	300	85	2,100
475	215	55	1,000

Calcium Distribution. The purpose of the carbonator and second clarifier (see flow sheet, Figure 1) is to recover lime which remains soluble in the discharge from the first clarifier. It is relevant, therefore, to ask how much lime is available for recovery and how much is precipitated in the first clarifier.

Calcium analyses were made on the supernatant liquid from a number of the jar tests described above. Calcium data from the tests mentioned above in reference to Figure 2 and Figure 3 are presented in Tables 4 and 5. These data show that the percentage of the calcium precipitated in the reaction is increased when pulping waste liquors are added to the effluent at any level of calcium dosage. Further, they indicate that as the lime dosage is increased, the percentage of the calcium precipitated also increases. In the proposed range of lime dosage, roughly half of the calcium will remain in the effluent leaving the first clarifier.

In any interpretation of the above data, it should be noted that the soluble carbonate content of the effluent will react with lime to form calcium carbonate, thus reducing the $Ca(OH)_2$ concentration available to promote color-reduction mechanisms. Since the carbonate content before lime addition was not determined, and carbon dioxide absorption was not rigorously avoided during the jar-test procedures, conclusions drawn from the data should not be over-extended.

Despite the reservations just stated, one clear conclusion is evident (in addition to the approximate fraction of lime in solution): the color-precipitation reaction is not strictly stoichiometric with respect to the calcium removed from solution.

TABLE 4

EFFECT OF TIME AND LIME DOSAGE ON CALCIUM PRECIPITATION

Calcium in Supernate, mg/l as Ca

Time Min.	Lime Dosage, mg/l $Ca(OH)_2$ 500	1,000	1,500	2,000
0	132	296	376	256
15	148	328	416	500
30	168	316	434	460
60	260	292	432	528
90	138	320	400	468
Average as $Ca(OH)_2$	312	574	762	892
% of Ca Precipitated	37.6	42.6	49.2	55.4

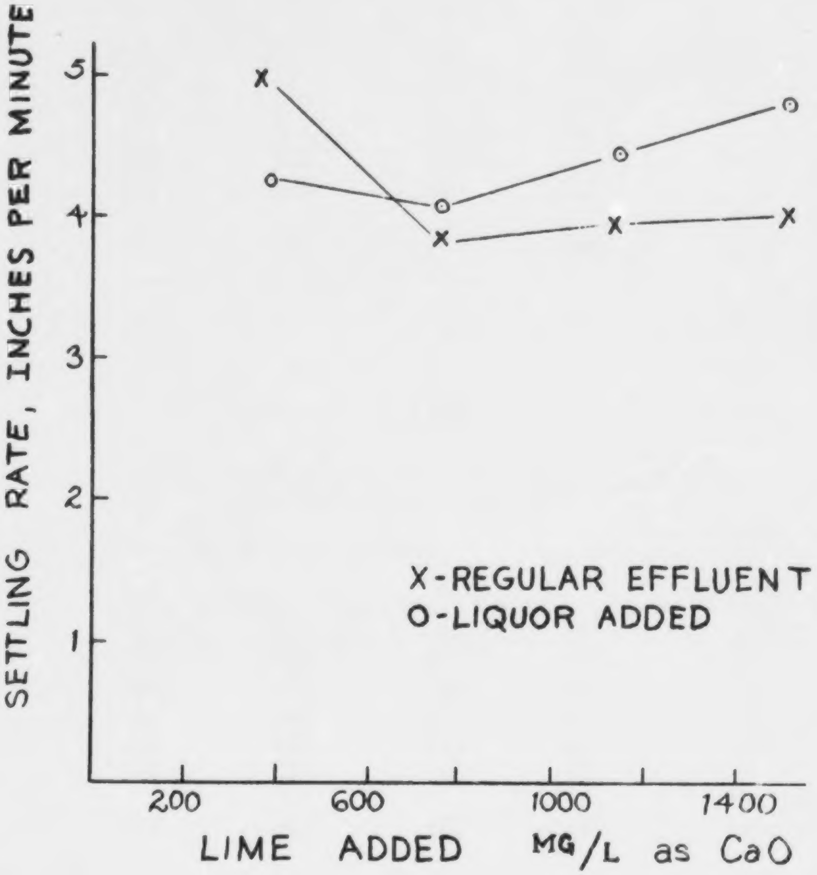
TABLE 5

EFFECT OF TIME AND LIME DOSAGE ON CALCIUM PRECIPITATION (WITH ADDED LIQUOR)

Calcium in Supernate, mg/l as Ca

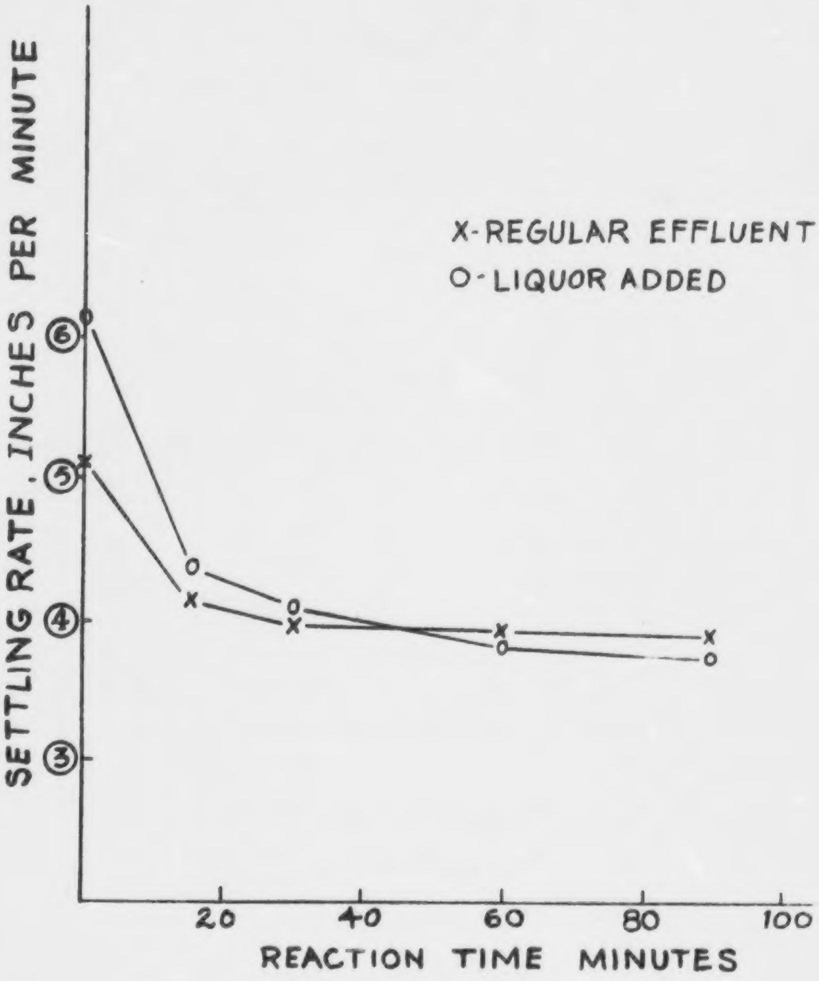
Time Min.	Lime Dosage, mg/l $Ca(OH)_2$ 500	1,000	1,500	2,000
0	116	228	280	364
15	136	260	324	400
30	104	272	368	456
60	148	244	340	484
90	104	296	412	452
Average as $Ca(OH)_2$	226	481	638	797
% of Ca Precipitated	54.8	51.9	57.5	60.2

Settling of Color Sludge. The effluent sample treatments used to generate the data for Figures 2 and 3 were also used to determine settling rates. The settling rate information is presented in Figure 4 and Figure 5. All of these rates were greater than 3.5 inches per minute. On this basis, it appeared that earlier projections of a clarifier rise rate of 1.0 g.p.m./ft.^{2} were amply justified. The adequacy of this design rate was also confirmed by observation of the pilot plant operation.



EFFECT OF LIME DOSE ON SETTLING

Figure 4



EFFECT [illegible] REACTION TIME ON SETTLING

Figure 5

Statistical Study. A half-replicate of a 2^5 factorial experiment was run to test the effects of reaction time, lime dosage, temperature, the influence of brown liquor, and the influence of black liquor. The response measured was color reduction. The data are presented in Table 6.

TABLE 6

FIVE - VARIABLE STATISTICAL STUDY OF PROCESS

Temp. °F.	$Ca(OH)_2$ mg/l	Time, Min.	Black Liquor	Brown Liquor	% Color Reduction
100	800	30	N	3N	44.4
120	800	30	N	N	63.2
100	1,600	30	N	N	80.0
120	1,600	30	N	3N	72.8
100	800	90	N	N	56.8
120	800	90	N	3N	55.0
100	1,600	90	N	3N	32.8
120	1,600	90	N	N	92.0
100	800	30	3N	N	84.6
120	800	30	3N	3N	74.1
100	1,600	30	3N	3N	80.2
120	1,600	30	3N	N	86.3
100	800	90	3N	3N	61.2
120	800	90	3N	N	75.4
100	1,600	90	3N	N	89.7
120	1,600	90	3N	3N	85.8

The analysis was done by a statistical specialist in Continental Can Company's Central Packaging Research Laboratory in Chicago. She reported the following results:

1. There was no detectable effect of temperature on time of reaction.
2. Increased lime addition improved the efficiency of color removal (95% confidence).
3. As the amount of black liquor was increased, the efficiency of color removal was increased (90% confidence).
4. As the amount of brown liquor was increased, the efficiency of color removal decreased (99% confidence).

The last two conclusions confirm strongly the earlier observations that the color of black liquor is much more easily removed than the color of brown liquor.

Sludge Quantities. The amount of sludge from the color precipitation (first) clarifier must be known, to insure provision of adequate dewatering and kiln capacities. Although estimates had been made on the basis of theoretical treatment of data already obtained, further data were obtained by a direct empirical approach. This was done using 50-gallon batches of effluent to which known amounts of lime were added. The sludge was allowed to settle and separate, and a dry weight was determined. The results are shown in Table 7. It is seen that the amount of sludge is roughly proportional to the amount of lime added, and that at a lime dosage of 1,200 mg/l, about 5 tons of sludge can be expected from each million gallons of effluent. The amount roughly doubles when the lime dosage is doubled. From this information normal sludge solids were projected on the basis of the 1,200 mg/l data, and peak loads on the basis of 2,400 mg/l level.

Obviously, sludge solids values will be affected by variations in fiber content of the effluent.

TABLE 7

ESTIMATED AMOUNT OF COLOR SLUDGE

Sample Date	CaO, mg/l	Sludge, Tons Per Day
11/06	1,200	55
11/06	1,200	52
11/07	1,200	59
11/07	1,200	59
11/07	1,200	62
11/09	2,400	105
11/09	2,400	138
11/09	2,400	122
11/10	2,400	118
11/10	2,400	120
11/11	2,400	135
11/11	2,400	139
11/14	1,200	78

Trials made with 50-gallon samples.
Calculated to total effluent of 12 million gallons per day.

Thickening of Color Sludge. Removal of as much water as possible from the color precipitation sludge prior to filtration or centrifugation is advantageous because it will reduce the design size of these expensive devices. Moreover, if filtration is integrated with that of causticizing sludge, excessive filtrate volume would increase soda losses in proportion to the amount of filtrate not usable as weak wash.

The response of color clarifier sludges to the action of a picket-type thickener was next studied. If such a thickener should prove effective, it was felt that either the sludge could be fed to a separate thickener, or the rake mechanism of the first clarifier could be designed to produce a thickening action. The second of these alternatives seemed the more attractive, since it would be less expensive. Suppliers were approached at this time to solicit suggestions concerning suitable mechanisms for this purpose.

In thickening sludges of this type, the critical dimension is the area of the sludge blanket, rather than the depth. "Unit area" (U.A.) is defined as the area in square feet required to thicken a sludge flow of one ton per day from a given starting consistency to a desired final consistency. The units of "U.A." are square feet per ton per day.

A series of laboratory thickening tests was run using sludge produced in the 2 g.p.m. pilot plant. One group was run using a lime dosage of 2,400 mg/l CaO, and another using 1,200 mg/l lime. Resultant unit areas required to achieve 10% solids with 2,400 mg/l group are shown in Figure 6. It can be seen that to thicken from 2% to 10% solids will require a unit area of about 40 square feet per ton per day. Since it has been estimated that there would be about 10 tons per day of sludge produced at this lime addition rate, a total thickening area of 400 square feet per daily million gallons effluent would be necessary. This is approximately half the area which would be provided by the most rapid clarifier rise rate which was contemplated.

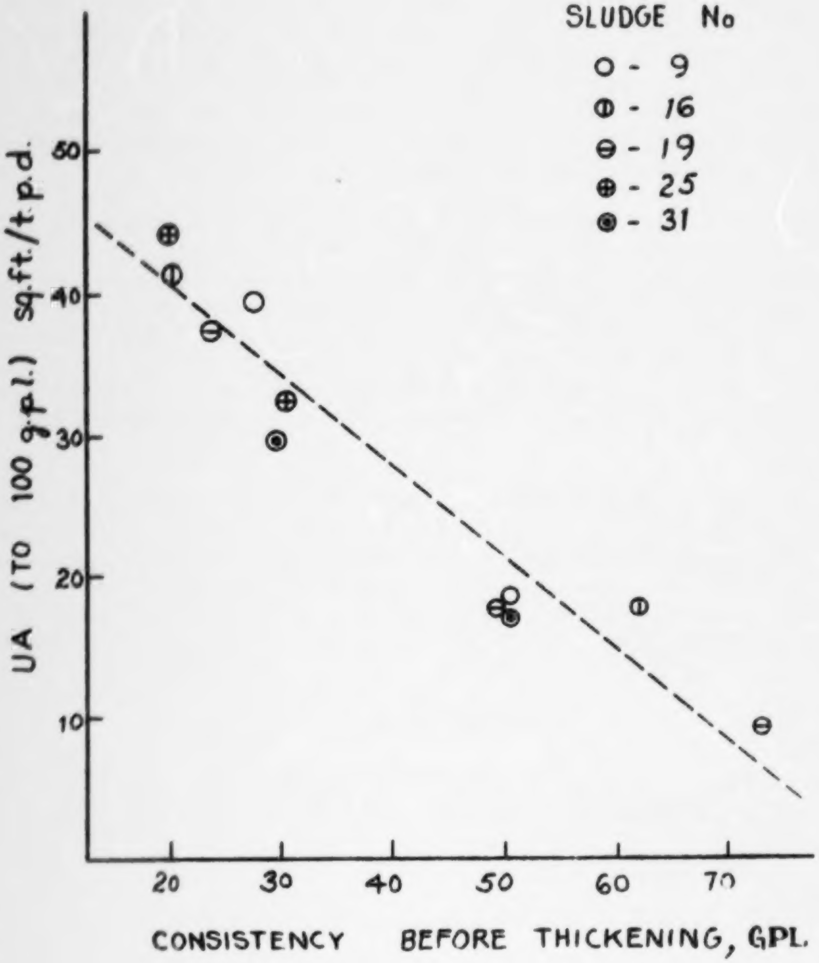
The analysis of the 1,200 mg/l lime addition group did not yield such neat figures. The reason for the scatter was not determined with certainty, but the data still appeared useful. (Variable fiber content may have been involved, and variable carbonate content would affect the active lime concentration, especially at the lower lime dose.) The data are shown in Table 8. The highest unit area shown is 154 square feet per ton per day. Since 5 tons of sludge per million gallons are estimated, a total thickening area 770 square feet per daily million gallons is required for this worst case. (A median value around 60 square feet per ton per day is indicated.) A capability prediction of 10% solids content for thickened sludge seemed reasonable.

TABLE 8

COLOR SLUDGE THICKENING DATA

Sludge Number	Initial Solids %	Unit Area Required*
5	2.2	67
5	4.4	52
6	3.2	45
6	6.3	34
8	2.9	47
8	5.8	27
24	2.1	99
25	1.9	154
35	2.0	104
35	1.0	114

*Unit Area (U.A.), sq. ft./t.p.d. to obtain 10% final solids. Effluent treated with 1,200 mg/l CaO, except No. 5 treated with 800 mg/l.



UNIT AREA FOR SLUDGE THICKENING	Figure 6

The calculations used to arrive at the conclusions in Table 8 were based upon the technique discussed by Rich (17). The data were also analyzed by a staff specialist of a major manufacturer of clarification equipment, and he arrived at essentially the same conclusions.

Sludge Filtration. A program of study was directed to the filterability of mixtures of primary clarifier sludge and causticizing lime mud. Mixtures of lime mud and effluent sludge were subjected to filter leaf tests. The filtering cycle used employed half of the cycle time for filtration, a quarter of the time for drying and a quarter of the time "dead." Filter leaf tests usually were made using 20, 30, and 40-second filtering durations. When other times were used, they are noted.

The importance of the relative amount of lime mud and sludge is illustrated by the data in Table 9. For this series, a quantity of effluent was treated with 800 mg/l CaO. The resulting sludge then was mixed with several amounts of lime mud taken at about 35% solids from the mill's mud washer. Lime mud alone will filter at the rate of 300 - 350 pounds per hour per square foot. Examination of the data in the table shows that when mixed with sludge, the filtering rates are considerably slower and that the rate is strongly influenced by the proportion of lime mud to sludge.

For most of the testing program, 50-gallon batches of effluent were treated with appropriate amounts of lime, and the sludge produced was mixed with a proportional amount of lime mud. In practice, the sludge from 50 gallons was mixed with 2.5 pounds (O.D. basis) of lime mud. This mixture represents the conditions of a mill using 175 tons per day of CaO for causticizing, and treating 12 million gallons of effluent daily.

TABLE 9

EFFECT ON FILTERING RATE OF
RATIO OF MUD SOLIDS TO SLUDGE SOLIDS

Mud/Sludge Ratio*	Filtration Rate**
3.0	29.1
4.9	54.2
7.0	95.2

*Lbs. mud solids added to sludge from 100 gallons effluent.
**(Lb.)/(Hr.)(sq.ft.), at 30 second formation time.

Some work was done to demonstrate the effect of the solids content of the mud-sludge mixture on the filtration rate. Two sets of data illustrating the effect are shown in Table 10. It is apparent that as the solids content is increased, the filtration rate is improved, and it is concluded that an effort should be made to provide as much mechanical dewatering capacity as possible in the system prior to filtration in order to minimize the work required by the filters. Picket thickeners would seem suitable devices for this purpose. These data also illustrate the marked improvement in filtration rate brought about by using a faster cycle. The negative aspect of this expedient is the thinner cake produced; if a belt filter were required, separation of cake from the filtering belt would become a problem if the cake became too thin and flexible.

TABLE 10

EFFECT OF SLUDGE CONCENTRATION ON FILTRATION RATE

Day	Solids Before Filtration, %	Formation Time Seconds	Formation Rate (lb.)/(Hr.)(sq. ft.)
10/10	10.0	18	44.0
10/10	10.0	36	65.5
10/10	14.3	18	59.9
10/10	14.3	36	82.2
10/17	9.3	30	73.4
10/17	9.3	45	59.0
10/17	9.3	60	51.0
10/17	12.8	30	114.0
10/17	12.8	45	91.7
10/17	12.8	60	82.0

There are presented in Table 11 average sludge/mud filtration rates for the days indicated. The data are arranged in order of increasing filterability, rather than in chronological order. Where available, the solids content of the mixture prior to filtration is shown. These data are applicable for a drum at 50% submergence, using a 30-second formation time. On the average, if a 20-second formation time were used, the filtration rates would be about 20% higher. It should be noted, too, that the initial solids content of the sludge mixtures average about 15%. It should also be remarked that thickening studies indicate the possibility that the solids content of material going to the filter might be as high as 25%, in which case higher filtration rates may be expected. It is apparent from the data that there is considerable day-to-day variation in the filtration rates. The reason for this variation was not known; perhaps it is related to the amount of fiber fines in the white water. Because of this variability, and because experimental data may never reveal the full range of possible conditions, considerable judgment must be exercised in sizing a filter for this application.

In Table 11 are also presented moisture data on the filter cake after filtration. An average of 48% solids is noted. When lime mud alone is filtered by the same technique, the filter cake is at about 60% solids. It is felt that in practice, the solids content of the mud/sludge cake will be about 50 - 52%, which is slightly higher than the laboratory data indicate, but still considerably lower than normal lime mud.

Carbonation and Carbonate Settling. Recovery of soluble calcium after the initial clarification next received some study. The flow scheme in-

TABLE 11

AVERAGE FILTRATION RATES

Day	Mud/Sludge Mix % Solids	Filtration Rate, (lb.)/(hr.)(sq. ft.)	Filter Cake, % Solids
10/19	----	31	48
10/21	----	32	45
10/14	14.0	33	--
10/22	13.0	49	45
10/16	----	54	52
10/20	15.3	57	50
11/10	----	58	50
11/07	15.4	59-1/2	47
11/09	----	64	49
10/23	----	66	46
11/06	10.4	67-1/2	49
11/09	12.2	71	47
11/08	14.5	72	47
11/06	19.1	103	48
10/18	----	106	45
10/17	----	123	50
		Average	48%

Formation time - 30 seconds

volves reacting the soluble calcium hydroxide with the carbon dioxide from lime kiln stack gas; the resulting calcium carbonate is then removed by sedimentation, dewatered, and calcined. The most important questions recognized at this point were:

1. Design of gas-liquid transfer system to effect reaction of the carbon dioxide.
2. Determination of the restraints establishing the extent of reaction to provide optimum recovery of calcium.
3. Mechanical considerations leading to most rapid and complete sedimentation of the calcium carbonate precipitate.

The first question seemed reasonable amenable to established chemical engineering knowledge and methods. Calcium carbonate scale formation and foam generation were recognized as problems.

The pH of the solution was taken as the readily measurable parameter by which to measure progress of the reactions:

$$Ca(OH) + CO_2 \longrightarrow CaCO_3 + H_2O$$

$$CaCO_3 + H_2O + CO_2 \longrightarrow CaH_2(CO_3)_2 \text{(Calcium bicarbonate)}$$

The pilot plant carbonator was run at different carbon dioxide flow rates to vary the pH of the carbonated effluent leaving. The carbonated effluent was centrifuged and an analysis made for dissolved calcium. The results are presented in Figure 7. They show a minimum dissolved calcium content at about pH 9.6. At lower pH levels, the dissolved calcium increases because of bicarbonate formation. Generally, low levels of dissolved calcium were observed over the pH range 8.5 - 10.5.

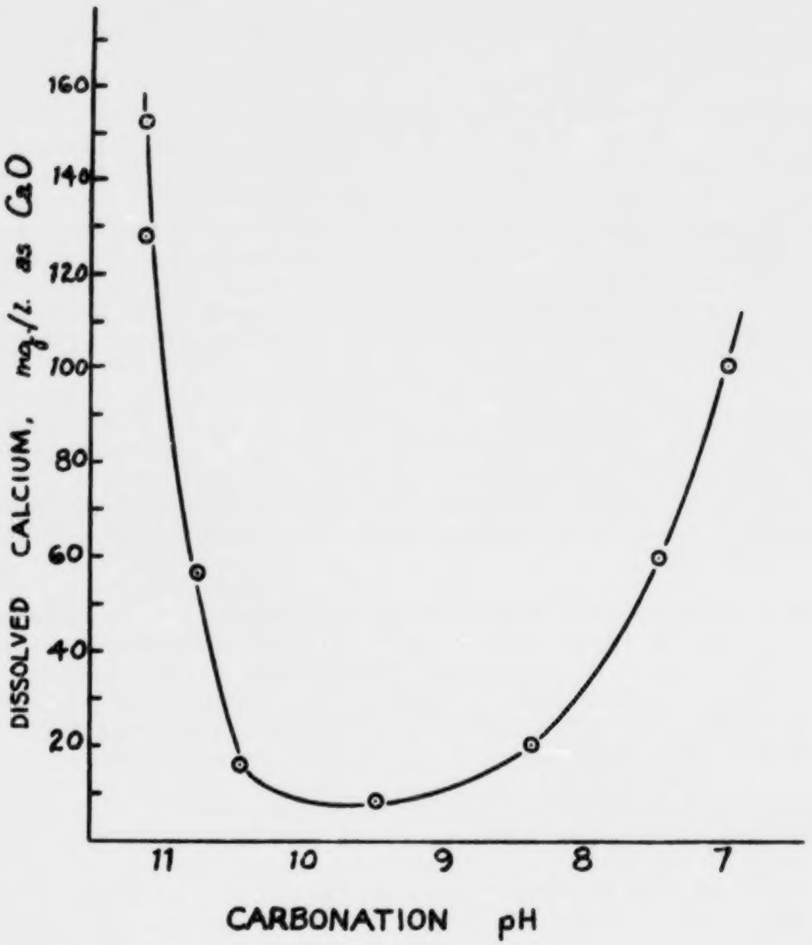
These pH values are notably lower than those reported for the "massive lime" study, (9) where a minimum solubility was indicated above pH 11. It is probable the difference is due to the fact the NCSI study dealt with bleach extraction wastes high in sodium alkali. In the presence of dissolved lime, this alkali is virtually all NaOH. Carbonation yields carbonate, and then bicarbonate -- both soluble. The carbonate from this source enters into the solubility product which determines solubility of calcium carbonate. It is to be expected, therefore, that the pH for minimum $CaCO_3$ solubility will vary with soda alkalinity of the starting effluent.

It was noted in the NCSI research that there seemed to be an optimum pH at which the calcium carbonate flocculated to provide good settling. Numerous observations were made in the present study, both of jar test and pilot plant runs; none led to a reproducible flocculating condition. There were scattered, transient evidences that raised hopes briefly, however.

The pilot plant system was operated to seek information on recovery at the post-carbonation clarification step. The Hodge pilot plant clarifier had severe limitations for this purpose. There was almost no opportunity for floc formation, and mean retention time was less than 20 minutes. The total up-flow path was only about two feet, and the calculated rise rate was about 1.0 g.p.m. per square foot.

The pilot plant was operated using a lime slurry pumping rate a little above 1,000 mg/l CaO while carbonation was carried out at three different pH levels. The amount of calcium entering and leaving the clarifier was measured, so that the clarifier efficiency could be measured. The results are shown in Table 12. About 70% of the calcium entering the clarifier was recovered; the best result was obtained at pH of 8.5, but the spread of results was small enough to raise a question whether they validly indicate a significantly preferable pH.

Considering the crude design of the pilot plant clarifier, this efficiency didn't seem bad. Since a similar process in municipal softening systems is operated routinely without difficulty, it was felt that 90% recovery in a properly designed, full-scale unit was not an unrealistic expectation. Because more than half the feed calcium would be recovered in the first clarifier, the indicated overall recovery would amount to about 95%.



EFFECT OF CARBONATION pH ON DISSOLVED CALCIUM

Figure 7

TABLE 12

EFFICIENCY OF PILOT PLANT CARBONATE CLARIFIER

Carbonation pH	Efficiency %
10.5	66.6
9.5	67.6
8.5	70.6

The calcium concentration to clarifier = 554 mg/l as CaO.

Treatment Suitability of Various Wastes. Samples from various branch sewers of the mill effluent system were collected to evaluate their relative color contributions and to provide a basis for determining which should not be (or need not be) treated by the proposed system. It was concluded that effluent from bag plant, power plant, water treatment and causticizing areas need not be treated. Talloil plant sewer waste was judged undesirable because of the occasional release of a batch of sodium sulfate brine. At high sulfate concentration, calcium sulfate will be precipitated, reducing the calcium hydroxide available for color precipitation, and contaminating the kiln product when the sludge is calcined. If a holding pond were available to distribute the impact over a suitable dilution of other effluent, a different choice might be made.

During examination of the various component subdivisions of sewer flow, it was noted that no tests were recorded in which the raw waste was primarily from a waste liquor source not modified by papermaking processes. Samples from the sewer of the digester and washer area were treated at two lime dosage levels. The results are shown in Table 13. They show that, on these generally dark samples, the treatment was able to remove in excess of 80% of the color, even at relatively low levels of addition.

TABLE 13

EFFICIENCY OF COLOR REMOVAL FROM PULPING AREA EFFLUENT

			Color Removal Efficiency, %	
Day	NSSC*	Effl. Color, APHA Units	Lime Added 760 mg/l	Lime Added 1,520 mg/l
11/08	Yes	2,813	85.8	86.7
11/10	No	1,500	66.7	91.7
11/11	Yes	3,500	71.4	78.6
11/12	No	1,950	76.4	----
11/13	No	3,190	81.8	91.9

*On "Yes" days, about 30% of pulp produced was NSSC (remainder, kraft).

Centrifuge Dewatering. Although the accumulated data indicated that an acceptable dewatering of color sludge could be accomplished by mixing with causticizing mud washer underflow and filtering with a belt filter, two disadvantages were recognized:

1. Soda removal from causticizing mud would be hindered, resulting in higher soda input to the kiln.
2. The belt filter requires closer attention than conventional, precoat-type mud filters.

Solid-bowl centrifuge dewatering of combined sludges had been considered earlier as an alternative to belt filtration, but was deemed unfeasible because of a tendency to leave lighter portions of the solids (e.g., precipitated color bodies) in the liquid discharge. Separate centrifuging of the first clarifier sludge had not received much serious consideration because of the low solids levels obtained in laboratory trials.

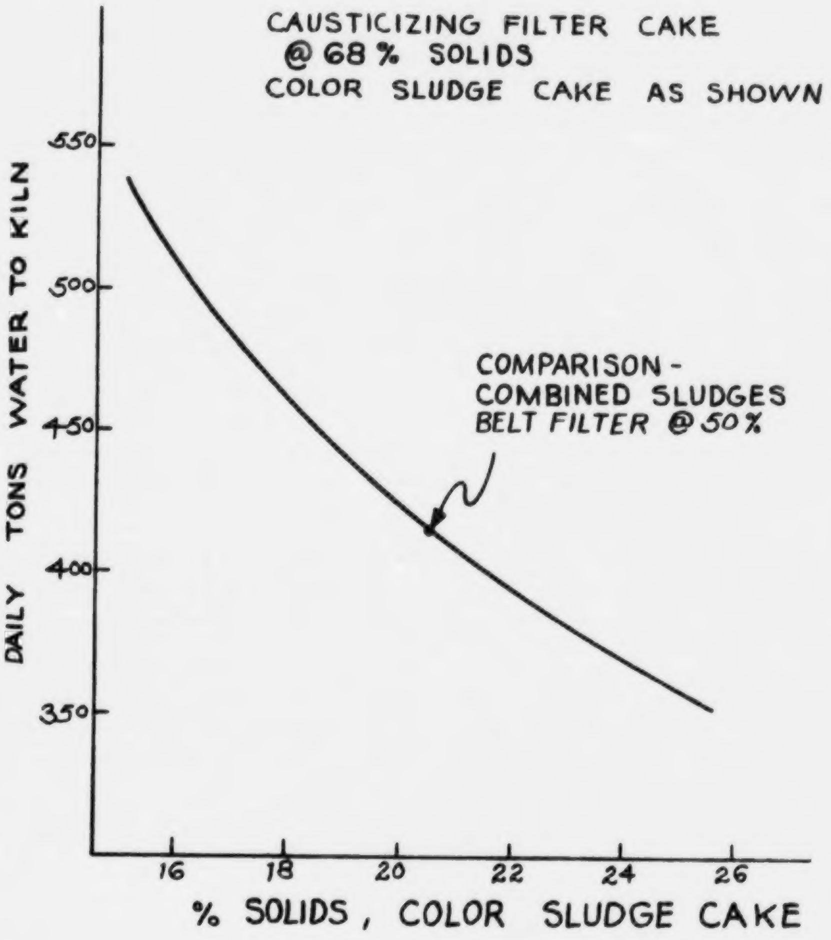
A fresh examination of process material balances, as well as the promise of better adaptability to remote operation control, led to further consideration of this technique. It became apparent that, since causticizing mud is commonly discharged at 65% solids from a conventional precoat filter, the separate concentration of first clarifier sludge to slightly over 20% solids would be comparable to the belt filter dewatering of combined sludges to 50% solids. Figure 8 shows the effect of increased sludge dewatering. Pilot-plant samples of sludge were prepared for testing by centrifuge manufacturers.

Two 5-gallon samples were tested by one manufacturer in specialized test apparatus with discouraging results. Their data indicated an assured solids concentration of only 15%, with appreciable solids carry-over in the centrate liquid.

A 55-gallon sample sent to another company was processed through a very small commercial type solid-bowl continuous centrifuge with quite promising results. Discharge cake solids contents of 20 - 25% were achieved; solids removal as high as 98% was found feasible by use of a flocculant. The data from this series of trial runs appear in Tables 14 and 15. Arrangements were made to rent a similar centrifuge both to conduct further tests and to prepare sludge for other experimental purposes.

A rented centrifuge was delivered to the mill, where pilot lots of sludge had been prepared for processing. The Sharples Model 600 centrifuge has a normal volumetric throughput maximum of about 6 gallons per minute. Standard rotational speed is 5,000 r.p.m., representing a relative centrifugal force of about 2,100 times gravity. Operating variables include:

1. Pond depth.
2. Speed differential of internal conveyor screw.



EFFECT OF COLOR SLUDGE DEWATERING

Figure 8

3. Feed slurry solids concentration.

4. Feed slurry flow rate.

5. Additives to aid flocculation and drainage.

The Model 600 is a small unit -- a choice made necessary by the difficulty in preparing large quantities of representative sludge. The inlet ports are small, and a small feed pump is required. As a result of these dimensions, (together with the fibrous nature of the sludge) high sludge concentrations were difficult to pump, and flow rate was difficult to adjust.

TABLE 14

CENTRIFUGE TRIALS (FACTORY)

Test No.	1	2	3	4	5	6	7	8
Feed:								
Code	A	A	A	A	A	B	B	B
Rate, #/Hr.	322	736	1315	1411	2400	432	1171	2215
% Insols.	3.25	3.25	3.25	3.25	3.25	2.93	2.93	2.93
Insols, #/Hr.	10.4	23.9	42.6	45.9	78.0	12.7	34.3	65.0
Centrate:								
Rate, #/Hr.	240	618	1120	1210	2160	390	1065	2040
% Insols.	.005	.027	.157	.189	.695	.088	.387	.600
Insols, #/Hr.	0.01	0.17	1.80	2.30	15.0	0.34	4.10	12.3
Cake:								
Rate, #/Hr.	82	118	195	201	240	42	106	175
% Water	87.2	80.0	79.1	78.2	73.8	70.5	71.5	69.7
Water, #/Hr.	71.6	94.3	154	157	177	29.6	75.8	122
% of Solids Unsedimented	0.10	0.71	4.20	5.00	19.2	2.70	12.0	19.0
Flocculant:								
Sol'n, gph	zero →							
Dose, #/T, s/s								
RCF x g.	2100 →							
Pond	3 →							
Convr. diff.	50rpm →					20rpm →		

TABLE 15

CENTRIFUGE TRIALS (FACTORY)

Test No.	9	10	11	12	13	14
Feed:						
Code	C	C	C	C	C	C
Rate, #/Hr.	1530	1530	1530	1530	3060	3060
% Insols.	2.55	2.55	2.55	2.55	2.55	2.55
Insols, #/Hr.	39.0	39.0	39.0	39.0	78.0	78.0
Centrate:						
Rate, #/Hr.	1347	1347	1336	1345	2770	2754
% Insols.	.037	.040	0.040	.177	.146	.088
Insols, #/Hr.	0.50	0.54	0.54	2.4	4.1	2.4
Cake:						
Rate, #/Hr.	183	183	194	185	290	306
% Water	79.0	79.5	80.2	80.2	74.5	75.3
Water, #/Hr.	145	145	156	148	216	230
% of Solids Unsedimented	1.3	1.4	1.4	6.2	5.3	3.1
Flocculant:						
Sol'n, gph	15.0	7.50	3.75	1.50	7.50	75.0
Dose, #/T, s/s	6.4	3.2	1.6	0.6	1.6	3.2
RCF x g	2100 →					
Pond	4 →					
Convr. diff., rpm	20 →					

As shown by Tables 16, 17, and 18, adjustment to comparatively low pond depth and differential speed resulted in cake densities in the range of 22 - 25%. Excellent recovery percentages were obtained with the use of a polymeric flocculant, and fairly satisfactory ones were achieved without such additives. It was considered likely that part-time use of a flocculant would suffice to prevent excessive buildup of "fines" in the system.

The potential advantage of separate sludge dewatering to such levels, as compared to the belt filter, was estimated by the following calculations:

Basis of calculations:

Causticizing mud solids	367 tons/day
Color clarifier solids	63 tons/day
Attainable % solids, conventional mud filtration	65 - 70%

TABLE 16

CENTRIFUGE TRIALS (MILL LAB)

Run No.	1	2	3	4	5	6	7	8	9	10	11
Centrifuge:											
Diff., rpm	51	51	51	51	51	51	51	51	51	51	51
Pond	3	3	3	3.5	3.5	3.5	3.5	3.5	3.5	4	4
Torque	1.0	1.5	1.2	0.5				1.0		1.2	
Effl. gpm	0.80	3.32	2.14	0.50	0.84	2.14	3.36	3.80	1.14	0.83	2.00
Solids, %:											
Feed	2.55	2.47	4.00	2.77	2.84	2.98	2.87	2.84	3.14	3.23	3.24
Cake	24.0	23.1	22.9	26.3	17.3	18.5	19.9	21.4	20.4	6.53	6.94
Effluent	0.15	0.49	0.74	.048	0.19	0.39	0.52	0.52	0.26	0.27	0.59
Recov., %	94.1	80.0	81.4	98.4	93.4	86.0	84.0	83.6	93.0	56.1	89.3
Flocculant, #/T, D.S.	none →										
RCF x g	2100 →										

TABLE 17

CENTRIFUGE TRIALS (MILL LAB)

Run No.	12	13	14	15	16	17	18	19	20	21	22	23
Centrifuge:												
Diff., rpm	51	51	51	51	51	51	51	51	51	51	51	51
Pond	4	3.5	3.5	3.5	3.5	3.5	3.5	3.5	3	3	3	3
Torque		1.5	1.2		2.0	1.2						
Effl. gpm	2.2	0.89	1.33	2.50	4.35	0.87	4.00	2.40	0.75	2.14	3.16	4.92
Solids, %:												
Feed	3.24	5.42	5.52	5.52	5.52	6.96	7.17	7.17	5.13	5.13	5.24	5.22
Cake	6.92	16.9	17.9	19.9	22.5	14.0	19.4	19.1	18.7	16.3	20.2	19.5
Effluent	0.57	0.46	0.61	1.31	1.48	2.18	2.00	2.00	0.26	0.91	0.95	1.35
Recov., %	89.7	94.2	92.2	81.7	78.5	73.1	0.5	80.5	96.4	86.8	86.7	80.0
Flocculant, #/T, D.S.	none	→										
RCF x g	2100	→										

TABLE 18

CENTRIFUGE TRIALS (MILL LAB)

Run No.	24	25	26	27	28	29	30	31	32	33	34
Centrifuge:											
Diff., rpm	30.6	30.6	30.6	30.6	30.6	30.6	30.6	30.6	30.6	30.6	30.6
Pond	3	3	3	3	3	3	3	3	3	3	3
Torque											
Effl. gpm	0.77	1.30	4.00	1.59	0.72	0.86	1.46	2.00	4.00	2.00	1.50
Solids, %:											
Feed	4.73	4.82	4.68	4.66	4.75	4.80	4.88	5.0	4.61	4.68	5.02
Cake	20.5	20.0	25.0	24.1	22.3	25.4	25.2	22.2	23.5	22.3	25.6
Effluent	0.56	0.31	0.86	0.32	.014	.029	.020	.047	.497	.017	.018
Recov., %	98.9	95.2	84.6	93.8	99.8	99.6	99.7	99.4	91.2	99.5	99.9
Flocculant, #/T, D.S.	none →				0.97	1.70	0.97	0.82	0.73	1.44	1.81
RCF x g	2100 →										

Attainable % solids, combined mud and sludge, belt filter	50%
Attainable % solids, color sludge, centrifuge	22 - 25%

Water Content:

430 tons combined sludge, 50% solids	430 tons water
(a) 367 tons caust. mud, 65% solids	197.6
(b) 367 tons caust. mud, 70% solids	157.3
(c) 63 tons color sludge, 22% solids	223.4
(d) 63 tons color sludge, 25% solids	189.0
Total, (a) plus (c)	421.0
(a) plus (d)	386.6
(b) plus (c)	380.7
(b) plus (d)	346.3

It can be seen that reduction in water to be evaporated in the lime kiln may range from 18,000 to 167,000 pounds daily.

It was felt that the belt filter would be much more difficult to adapt to remote control and minimum manpower requirements. Moreover, maintenance considerations appeared to favor the centrifuge over the belt filter. A flow sheet incorporating the centrifuge alternate is shown in Figure 9.

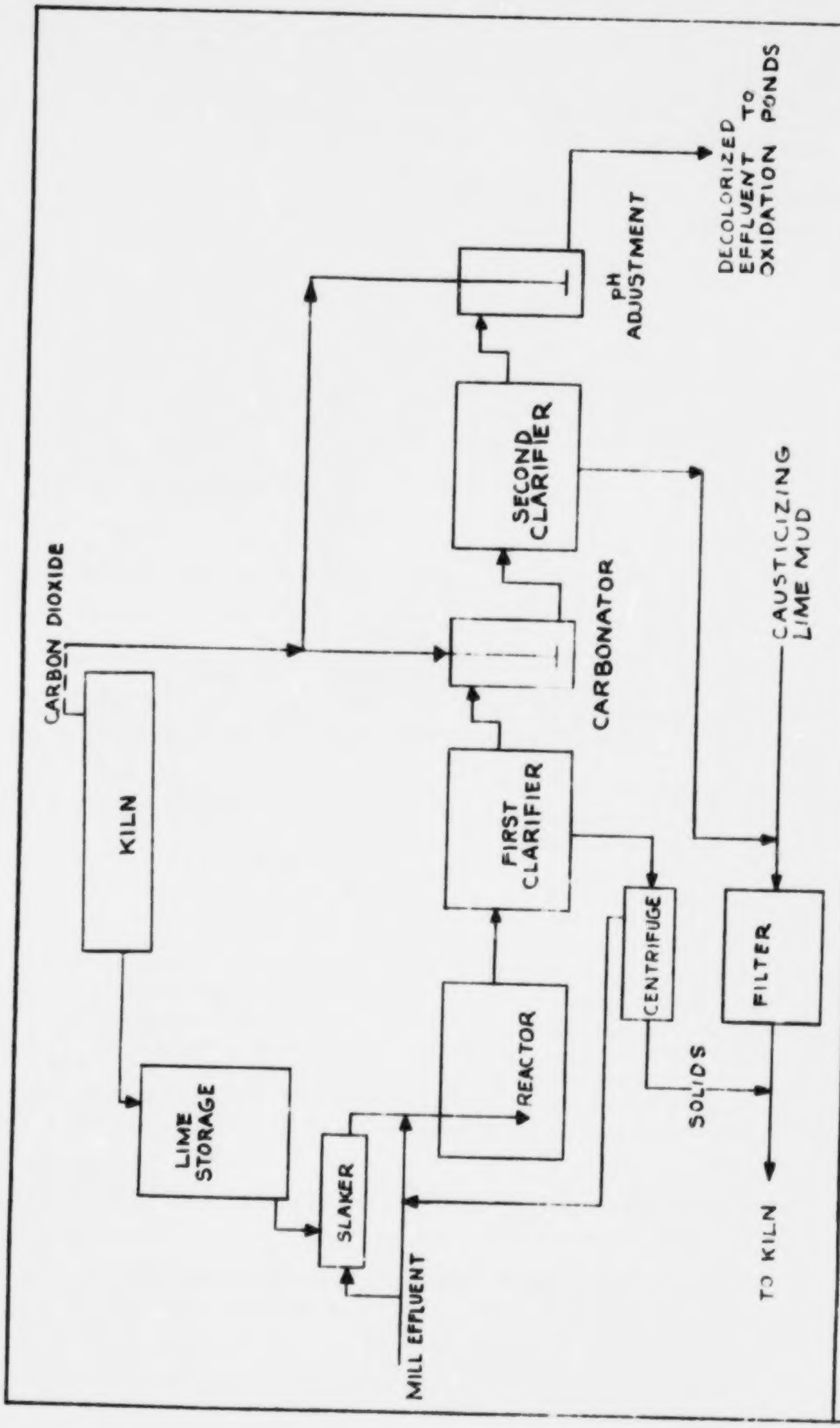
Sludge Calcining. During consultations with rotary kiln manufacturers concerning special design and operational considerations for sludge calcining, a warning was voiced. Engineers for one manufacturer noted that small increases in free lime content of lime kiln feed have caused serious problems of ball and ring formation in a number of kilns.

The question of whether the primary sludge of this process contains appreciable $Ca(OH)_2$ had received much attention. Solubility considerations would seem to cast doubt upon such occurrence, because the surrounding liquid is not saturated with calcium hydroxide. However, the sludge will provide hydroxyl ions to causticize a solution of sodium carbonate, and (as has been previously noted) the calcium content of the sludge does not follow a definite stoichiometric relationship to identifiable calcium-precipitating anions.

Discussions of the problem with engineers of two kiln manufacturers raised the following questions, among others:

1. Will the associated organic materials, co-precipitated with the calcium and hydroxyl ions, alter the effect, compared to calcium hydroxide?

2. Will the fiber content serve to reduce the balling and ringing tendency?


CARBON DIOXIDE
KILN
LIME STORAGE
SLAKER
MILL EFFLUENT
REACTOR
FIRST CLARIFIER
CENTRIFUGE
SOLIDS
TO KILN
FILTER
CARBONATOR
SECOND CLARIFIER
CAUSTICIZING LIME MUD
pH ADJUSTMENT
DECOLORIZED EFFLUENT TO OXIDATION PONDS

FLOW SHEET WITH CENTRIFUGE

Figure 9

3. Are there operational correctives which would allow successful calcining in the presence of free lime?

The first two questions attracted no firm opinions, and the third elicited disagreement among the experts.

Laboratory calcining had established the satisfactory chemical composition of lime made from the process sludge, but such procedures offered little guidance to physical behavior in a rotary kiln. Further conferences with a kiln maker dealt with the possibility of pilot-scale calcining as a means of resolving the questions. A pilot-plant kiln, 1' diameter by 10' long, was available which would require some 50 pounds per hour of feed (recausticizing sludge with proportional color-removal sludge). It was felt provision should be made to run up to 20 hours. With our small-scale facilities, and the problems of thickening, the accumulation of sufficient color-removal sludge was tedious, but the availability of the small, solid-bowl continuous centrifuge made it possible to concentrate it properly.

A quantity of the centrifuged sludge containing about 200 pounds of solids was mixed with causticizing sludge filter cake containing about 1,500 pounds of solids. This mixture was subsequently calcined, with encouraging results. This operation was conducted and observed by a group experienced in interpretation of miniature kiln performance. Their report was that a slight tendency toward ring formation was evident, but that the rings were rather fragile. Their opinion was that, in a comparatively large kiln, stable and troublesome rings were not likely.

Meanwhile, alternative flow sheet modifications were developed to eliminate the free lime before calcining, if such a course should become necessary. The first clarifier sludge would be carbonated before dewatering. In the case of centrifuge dewatering, the centrate might be used for lime mud washing if color release presented barriers to recycling to the clarifier.

Color Changes after Lime Treatment. As the project was approaching the point of major construction commitment, there were several warnings of "color reversion" or "color pick-up" possibilities. The warnings were investigated, and several experiments were conducted. The following conclusions were reached:

1. Colored substances are removed and incinerated by the process; this color cannot return.

2. Due to soda alkalinity, color can be extracted by effluent from woody and other plant materials in lagoons or streams.

3. Biological processes may increase or decrease color of effluent after lime treatment. Some quite varied observations were made, inviting further study for which time was not immediately available.

4. Although the effects noted in (2) and (3) might affect ultimate color effects in receiving waters, they do not negate the effects of the treatment.

Published Information. Work done in the bench and pilot phases of the project has been the subject of a published paper. (18) The theoretical considerations upon which the project conception was based has received support in studies directed by Dence and Luner. (19)

Plant Design

Before entering into specifics of unit designs, several basic decisions were necessary to establish the working flow sheet. It was determined that:

1. The sludge produced in the color precipitation step would be dewatered by a continuous, solid-bowl centrifuge. The operational and economic considerations bearing upon this choice have been noted previously in the discussion of the centrifugation studies.

2. A separate reaction vessel would not be used for color precipitation. Instead, lime would be added in the feed pipe to the clarifier center flocculation well, which would be designed for maximum reaction time.

3. A sewer conduit from the mill would deliver waste water to a lift station which would pump to a first clarifier which would be at such a level that gravity flow would suffice for further transportation of the main waste stream. There had been earlier hopes that gravity flow could be maintained throughout, but the only feasible sites presented unacceptable sewer problems.

4. To provide for a then-current need for an average daily capacity of 12 million gallons and a 50% reserve capability for peak rates and for possible changes, the principal system units would be designed for a maximum flow of 12,500 gallons per minute.

5. Provision would be made for removal of trash and grit from the waste water before adding lime. Previous experience with primary clarification of effluent had pointed out the frequent appearance of such foreign materials as sticks, rags, and pieces of string, rope, and gasketing. These had caused serious problems in a simple clarifier system; the complex system now proposed would be far more vulnerable to debris. Moreover, the centrifuge would present a special sensitivity to gritty, abrasive materials.

6. Use of lime, whether freshly purchased or re-calcined, would be common to causticizing and waste treatment, without separation or distinction.

7. Calcium carbonate sludge, obtained by carbonation of the first clarifier effluent, would be mixed and filtered with the recausticizing "lime mud", using conventional kraft mill equipment.

8. Carbonation for lime recovery would be accomplished by use of lime kiln stack gas after "scrubbing" for dust removal.

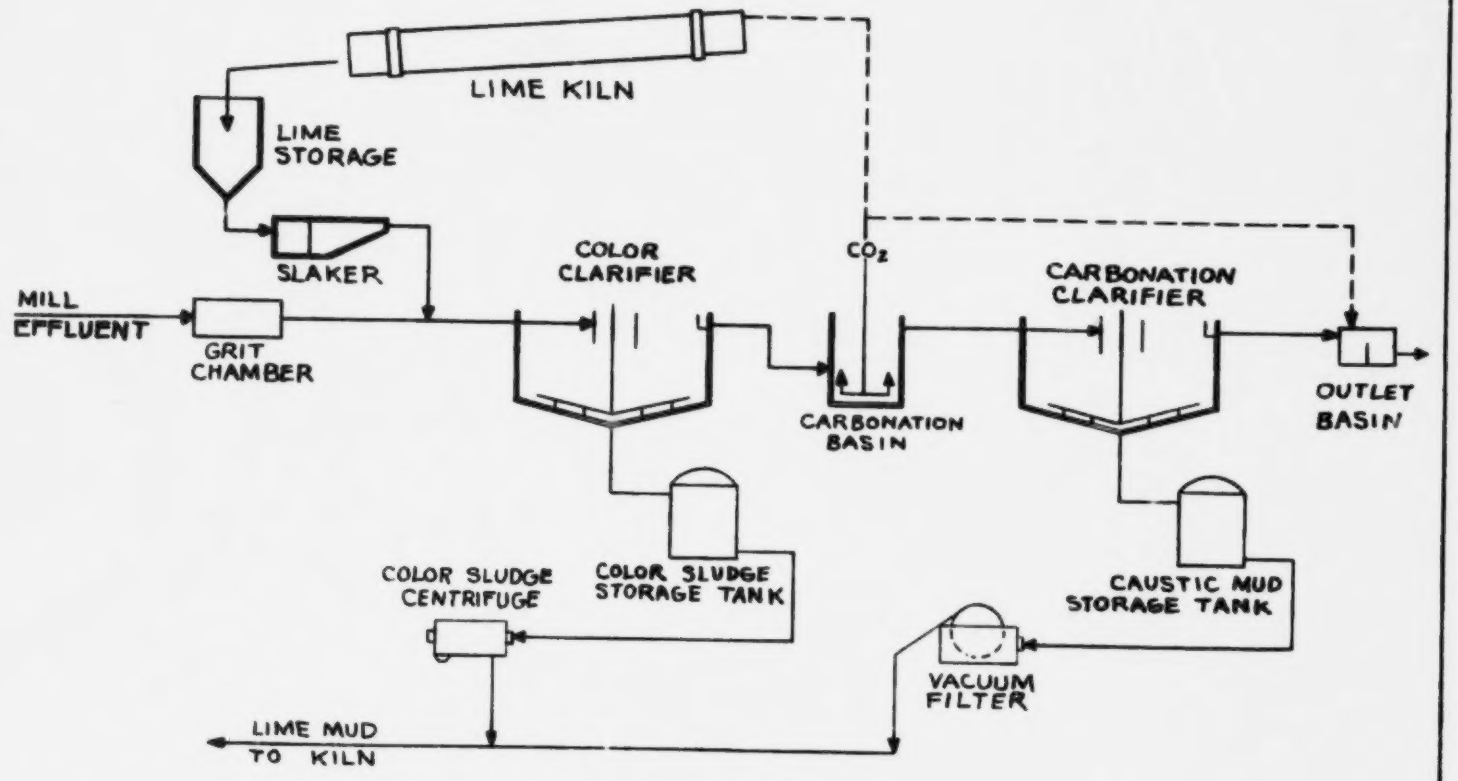
Upon reviewing these decisions, along with the laboratory and pilot plant work, the flow sheet was drawn in the form shown in Figure 10. As a basis for preparing detailed cost estimates and for soliciting bids, equipment needs were defined and specified. The following summary deals with equipment in process flow sequence, rather than in the chronological order of actual consideration.

Grit and Trash Removal. Earliest consideration of the removal of troublesome solids from the raw waste had envisioned a traveling bar screen followed by a small basin for collection of grit and other small, dense solids. Further consideration of specific materials found in the effluent soon revealed need for more complex design. It could be seen that a bar screen would not protect adequately against strings, bits of rope, or slender pieces of wood, all of which would then pass through pumps, control valves, and the entrance ports of a centrifuge. Although expected grit volume was not large, it was apparent that other materials, such as woody rejects from cleaners and screens, and water-logged bark would also settle and would comprise sufficient volume (at least occasionally) to demand a well-designed removal and disposal system. When a centrifuge of very high centrifugal force was chosen for sludge processing, it was noted that grit content of the sludge might be a prime factor in maintenance costs; obviously, effective grit removal assumed added importance.

After consideration of several devices, our choice finally settled upon a revolving disc screen. Specifications were prepared for a unit 14 feet in diameter with screen plates of Type 304 stainless steel, with 1/2 inch diameter holes. A solids-collection drag conveyor was required to gather, dewater, and deliver solids to a container, and a spray system was provided to flush debris from the screen onto the conveyor.

The grit settling capacity was calculated upon a minimum capability, at peak flow, to remove all grit above about 65 mesh. A chamber 10 feet wide (bottom sloped to 5-foot recessed channel) and 60 feet in effective length was provided, with a normal water depth of 7 feet 6 inches. Construction details include an overflow weir in the grit channel and a discharge weir after the disc screen, so proportioned that differential level across the disc screen cannot exceed about 18 inches. (Because of the large screen radius, drastic forces might otherwise develop in case of screen "blinding.")

The grit collector is designed to rake a bottom channel 5 feet wide, with buckets approximately 8 feet apart, and at a speed not exceeding 8 feet per minute. The mechanism is specified to be capable of pulling


LIME KILN
LIME STORAGE
SLAKER
MILL EFFLUENT
GRIT CHAMBER
COLOR CLARIFIER
CO_2
CARBONATION BASIN
CARBONATION CLARIFIER
OUTLET BASIN
COLOR SLUDGE CENTRIFUGE
COLOR SLUDGE STORAGE TANK
CAUSTIC MUD STORAGE TANK
VACUUM FILTER
LIME MUD TO KILN

FLOW SHEET FOR PLANT DESIGN

Figure 10

through a 9 inch grit accumulation at the bottom of the grit channel, dewatering and elevating the sediment to discharge into a container.

To minimize wear, neither the screening system nor the grit removal mechanism is operated constantly. Operation of the screen and its discharge conveyor are initiated whenever head loss across the screen exceeds 2 or 3 inches, and the ensuing timer-controlled cycle is set for a minimum of 3 minutes duration. The grit conveyor operates under an adjustable timer control, commonly set for 10 minutes running time each 60 - 120 minutes.

Lift Station. To collect accepted effluent from the disc screen, a pumping pit 18 feet in diameter by 17 feet deep was designed. Construction consisted of a circle of sheet piling, which was excavated deep enough for a ballasting bottom of mass concrete.

Pump specifications provide three pumps, with 16 feet tube length, each with capacity for 9,000 GPM at 30 feet head and 6,500 GPM at 45 feet head. These are intended to accommodate peak flow with two units running, thus allowing one as a spare in case of failure or routine maintenance of any unit.

The pumps, through check valves, discharge into a manifold connecting to a 24-inch pipe. It is in this pipe that lime slurry is mixed with the raw waste.

Lime Slurry System. Lime for the treatment process is withdrawn from a "day bin" which is shared with the kraft causticizing system. A screw conveyor with manually-controlled, variable-speed drive regulates lime flow into a fixed-speed transfer conveyor (also screw type) which delivers into the slaker inlet. The variable-speed conveyor has a screw 14" in diameter with a speed range of 1 - 3 RPM. The transfer conveyor screw is also 14 inches in diameter.

The lime slaker is a renovated unit originally used for kraft liquor causticizing. It has a reaction bowl 9 feet in diameter by 7 feet deep, supplied with a 15 HP paddle-blade mixer. The grit classifier is of the oscillating rake type, with a 2 HP motor.

The slaker overflows into a slurry tank 12 feet in diameter by 10 feet deep. The slurry is kept in suspension by a 7.5 HP vertical mixer.

Two slurry pumps (one a spare) are provided to deliver the lime suspension from the slurry tank to the raw waste lift pump discharge. The pumps are specified as rubber-lined (replaceable liners) for resistance to abrasion. The required performance capability is 180 GPM at 55 feet head, with a maximum pump speed of 1,200 RPM. The pumps discharge through a pipe about 1,100 feet long, and no flow throttling is provided. Water is supplied to the slurry tank to maintain a constant level; thus, the pump delivers full discharge pressure maintaining a velocity which should preclude deposition on the pipe walls. An automatic flushing (water) cycle is initiated whenever the pump is de-energized.

Waste Clarifiers. Since design criteria for both clarifiers were based on maximum rise rates of 1.0 GPM per square foot, it was possible to design them for identical internal mechanism. (In addition to maintenance advantages, it would be possible to assure one primary clarifier in case of failure of the other unit.)

The mechanisms were specified for a clarifier 135 feet in diameter, having 15 foot side water depth. A flocculation zone equivalent to a center well of 40 feet diameter was required. Thickening-type mechanisms capable of developing 7% (w/w) minimum (first clarifier) sludge consistency, based on pilot plant solids quality data, were specified. Since shear properties and maximum consistencies of both sludges represented first-of-a-kind design problems, it was determined that torque capabilities should be the highest competitively available in standard manufacture. This proved to be a continuous rating of 1,200,000 ft-lbs., with peak load design of 1,800,000 ft-lbs. Two long (full clarifier radius) and two short arms (25% minimum radius) were designed for sludge raking. Sludge is raked into an annular central well, within which full raking was required. A torque indicator, with transmitter for continuous, remote recording of rake torque, was included in the specification. Side or top entry of raw waste feed was specified; the final choice was for over-the-top entry, with feed pipe suspended below an access walkway bridge.

The clarifier basin designs were for concrete construction. Bottom slope to the center was one inch per foot. Clear liquid discharge at the upper periphery is through 66 equally-spaced, submerged orifices 6 inches in diameter. An integrally-cast, external, concrete collection trough, 3 feet wide and 6 feet deep, delivers the effluent to a discharge box.

Sludge withdrawal is accomplished through two stainless steel pipes cast in concrete beneath the clarifier floors. Because of the clarifier elevation, the color clarifier discharge pipes (8 inches nominal diameter) are essentially straight, and the ground-level pumps are only slightly above the elevation of the inner slope of the clarifier floor. The carbonation clarifier is slightly lower; the sludge pipes (6 inches diameter) lie parallel to the clarifier bottom and have slight bends up to pumps at ground level (clarifier water level is 12 feet above pump suctions).

Carbonation System. To provide for proper contacting of color clarifier effluent with kiln stack gas (source of carbon dioxide), a tank 30 feet in diameter by 12 feet deep was designed. Water from the color precipitation clarifier enters just below the water level of the carbonation tank and flows out near the bottom.

Proposals were considered for one, three, and four agitators. The final choice was for four 40 HP units with turbine-type impellers driven at 85 RPM. Considerations leading to this selection included; a more desirable subdivision of gas entrance points, ability to withstand temporary outage of a unit, and avoidance of sidewall baffles which would affect static pressures at entrance and exit openings.

It was considered probable that impellers of these agitators would accumulate deposits of calcium carbonate, and various expedients were considered for protection. Highly-polished surfaces, various alloys, elastomeric coatings, and repellent materials such as the fluorocarbon polymers, were among the proposals. None was found to give enough promise to justify the cost, and, since there seemed to be no serious corrosion problems, ordinary steel was specified.

Pilot plant experience and reports of related experience elsewhere had indicated that perforated-pipe distribution of carbonation gas would result in closure of the pipe openings by scale formation. It was also felt that gas pipes entering through the tank bottom would be subject to occasional back-up of water which might tend to be trapped, increasing back-pressure. (Required gas pressure is a substantial factor in cost of supplying carbonation gas.) The final design of the gas diffuser is sketched in Figure 11. Each diffuser discharges two 36-inch curtains of gas bubbles beneath one of the agitators. Stainless steel supply pipes, 6 inches in diameter, enter from the top of the tank. The diffusers consist of half-cylinders of 6 inch radius, 36 inches long, with 1-inch, 60° serrations on the long edges.

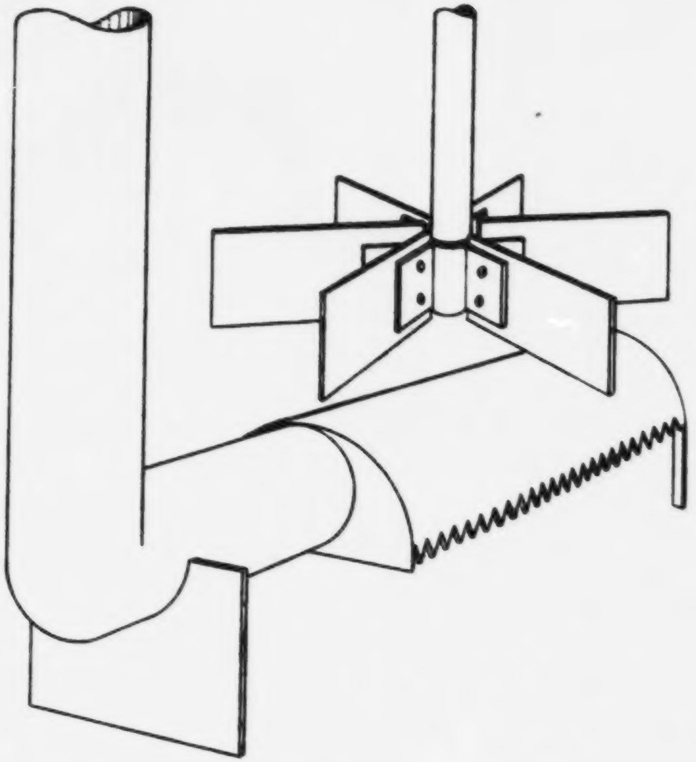
Outlet Basin. The carbonation clarifier discharges into a basin about 21 feet by 27 feet, and 12 feet deep. More kiln gas is added to adjust pH. Horizontal, perforated pipes are used for gas diffusion; as pH drops below 9.0, calcium carbonate solubility increases, so scaling should not be a serious problem. The basin also receives any overflow from the color clarifier or carbonator. From the basin, water can be discharged over adjustable weirs to either the secondary treatment system or to a holding pond. Provision is made for vertical pumps by means of which the "decolorized" water can be returned to the mill for selective re-use.

Sludge Pumps. To withdraw sludge from the color clarifier, specifications were based upon a sludge of 7% consistency in which fiber comprised 20% of the total solids weight (dry basis). Stated pump requirement was for a horizontal centrifugal pump capable of handling a 10% solids content, delivering 300 GPM at 170 feet total discharge head, with a maximum pump speed of 1,200 RPM.

For carbonation clarifier sludge, a sludge consistency of 10%, comprised essentially of calcium carbonate, was assumed. A capacity of 100 GPM at 120 feet total discharge head was required, with a maximum pump speed of 1,200 RPM.

Changes were made in both pumps after start-up, as will be explained later. The changes were dictated by the composition and properties of the sludges.

Color Sludge Storage. To receive sludge from the color clarifier, a 30



GAS DIFFUSER

Figure 11

foot diameter tank 20 feet deep was provided. The tank was identical to one specified for causticizing lime mud, and similar 15 HP agitation mechanism was provided. (Supplementary agitation was added later, as will be noted.) This tank provides surge capacity between the clarifier and the centrifuge.

Color Mud Pump. To deliver color sludge to the centrifuge, a rubber-lined centrifugal pump, similar to those for lime slurry, was specified. The required rate was 185 GPM at 80 feet discharge head.

Color Sludge Centrifuge. Considerations in centrifuge requirements have been discussed previously, especially as relates to dewatering capability. Specifications were based upon observations which indicated that optimum input consistency of sludge might be about 8%. Stated solids capacity was 70 tons per day. Automatic protective devices, to protect against torque overload and other probable sources of equipment hazard, were specified.

As a consequence of the required degree of dewatering, the only supplier offering to meet the specifications was one who offered a high "G-force" unit. Relative centrifugal force was 1,880, at a bowl speed of 2,300 RPM. The backdrive spindle of the planetary gear system rotating the internal conveyor was provided with an eddy-current electric brake which permits variation of conveyor differential speed by a ratio of two to one (approximately 25 to 50 RPM). The protective alarm system includes a signal to discontinue feed and/or provide flushing at a preset level of conveyor torque, and trip-out for low oil pressure, low oil flow (either bearing), high bearing temperature (either bearing), excessive torque, low brake-cooling water pressure, or high brake-cooling water outlet temperature. A fault detector display panel indicates trip-out cause.

The centrifuge solids are delivered by a screw conveyor into the lime kiln feed screw hopper, along with discharge cake from the kraft causticizing mud filter. Filtrate flows to the lift pump station to be reprocessed.

Carbonation Sludge Recovery. Piping was arranged to deliver the recovered calcium carbonate underflow from the carbonation clarifier to the causticizing "mud" storage tank. This would permit the combined calcium carbonate sludges to be dewatered by the lime mud filter, which was generously sized to provide the needed capacity to prepare the materials for calcining.

The Lime Kiln. Design of the lime kiln was based upon the expected requirements of the color removal system, combined with those for liquor recausticizing for about 600 tons of kraft pulp production. Kiln specifications required a capacity of 235 tons of lime per day at 85% available calcium oxide. Conditions for this performance included 45.2% sludge solids, and a total dry solids load, including fiber and recirculated dust load, of 422 tons per day. The indicated moisture load was well above the expected level, to insure adequate chain section provision. The kiln size was set at 290 feet by 12 feet inside shell diameter, with

6 inch refractory lining. The successful bidder provided a 58 foot chain system, weighing 130,700 pounds. Included was a venturi scrubber for stack gas, rated at 99% dust recovery, and an induced draft fan designed to provide a 25 inch (water) pressure, while handling 101,500 ACFM at 400°F.

Kiln Gas Blower. To supply lime kiln stack gas to the carbonation system, specifications were drawn for a compressor having a capacity of 4,000 ACFM at 160°F, at a discharge pressure of 10 psig. Gas is received from the discharge of the venturi scrubber system at a pressure of 1.0 atmosphere.

Since scrubber specifications indicated a $CaCO_3$ content of only 0.08% by weight, bids were offered by manufacturers of rotary lobe-type compressors, who represented their units as suitable. These units were priced lower than the "water-piston" type compressors, which the development engineers had considered most suitable for the service.

Attention is called to subsequent findings concerning both the equipment performance and the required gas volume.

Instrumentation. The needs for process instrumentation are largely influenced by three important considerations. First, the system is an added responsibility of the operator of another process, and the demands upon his time should be limited. Second, much of the equipment is relatively remote from the normal location of operating people. Third, the system is ponderous and slow with respect to most functions and responses. The last factor mitigates considerably the sophistication which might otherwise be necessitated by the first two.

Control of the grit conveyor and disc screen system by a timer and differential level sensor, respectively, has already been explained.

A level recorder in the control room, together with a "high level" annunciator light, inform the operator concerning lift station performance. The same level measurement is used to actuate a pump discharge valve so as to maintain a constant level in the pit, maintaining pace with the amount of effluent arriving.

Lime feed to the slaker is controlled by a manually-set speed control governing the feed conveyor; the speed is monitored by an indicating tachometer on the control panel. Water supply to the slaker is regulated by an indicating flow controller. Recording thermometers are provided for water input and slurry overflow from the slaker; temperature rise is a measure of calcium hydroxide concentration of the product slurry. (Conversion of one pound of CaO to $Ca(OH)_2$ releases 486 Btu. Thus, production of a 10% calcium hydroxide slurry is accompanied by a temperature rise of about 38°F.) (20)

The lime slurry tank is provided with a level indicator-controller. The controller actuates a water valve to maintain a predetermined level in the tank (usually about 6 feet). Low water level will activate a relay to shut off the slurry pump. (The slurry pumps are also interlocked to prevent function if all lift pumps are inactivated.) Upon trip-out of the slurry pump, a switch and timer provide flush-out of the slurry piping.

Torque exerted upon the rake of the color clarifier is recorded at the control panel. A separate limit switch is connected to an annuciator light to signal "high torque." Identical signals are provided at the carbonation clarifier, and torque for both units is recorded on the same 2-pen instrument.

At the carbonation tank, pH measurement is indicated locally and transmitted to a recorder-controller at the control panel. A signal from the controller actuates the control valve admitting kiln gas to the carbonation system. Both pH and valve position ("% open") are recorded on the 2-pen recorder.

Constant pressure in the kiln gas supply line, as well as protection (pressure relief) of the compressor, are provided by a pressure controller and vent valve. A control panel station permits adjustment of pressure and observation of valve position.

Control of pH at the outlet basin is facilitated by pH instrumentation substantially identical to that for the carbonation vessel.

Control and recording of sludge flow from each clarifier to storage is provided. In each system, a magnetic flow meter signal is recorded, and control action is supplied to a valve designed for precise flow control. Because of the plugging hazard inherent in the sludge properties, a parallel valve circuit with full pipe diameter, pneumatically-actuated plug valve is installed; this unit can be positioned by a manual control on the main panel.

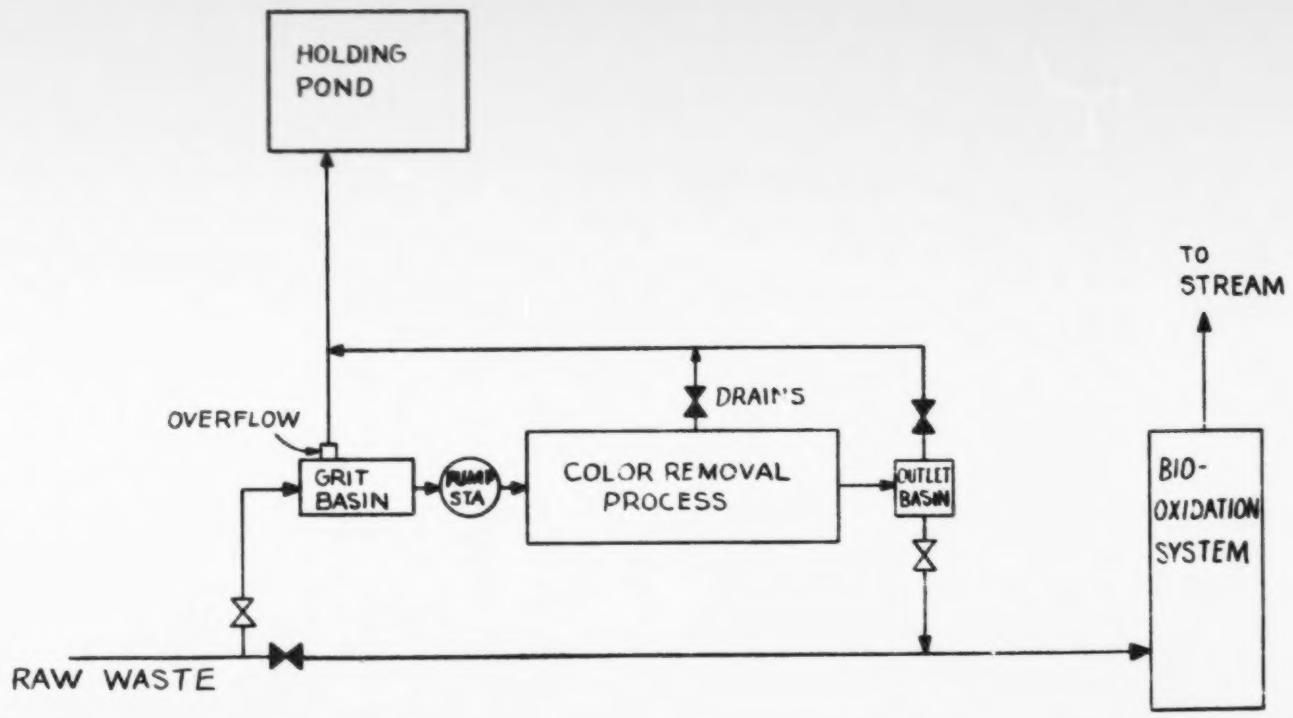
The storage tank for color clarifier sludge is equipped with level measurement. This measurement is recorded, and connections to the annunciator panel activate alarms to warn of high or low levels. A recorder to display torque loading of the tank agitator paddles was planned, but a suitable sensing element was not available.

Flow of sludge to the centrifuge is controlled and recorded by instrumentation at the main panel. To provide maximum uniformity of flow to the centrifuge, a concentrically constricted, elastomer-sleeved valve ("sphincter valve") was desired. Vendors were hesitant, lest the sludge prove too abrasive for this type valve, so a stainless steel V-ball unit was used.

Torque load required to drive the internal conveyor to the centrifuge is continuously recorded. The signal for this measurement is an electrical measurement of "percent excitation" of the eddy-current brake system previously ventioned. A manually-set percent-excitation value (commonly 60 - 65%) will serve to interrupt sludge feed and flush the centrifuge, and at

95% of brake rating, the centrifuge motor would trip out. Further details of protective devices have been noted in description of the centrifuge.

Emergency Provisions. To provide for short-term malfunctions of the system, including equipment failures or capacity overloads, as well as for shutdowns for changes, the arrangement shown in Figure 12 was designed. Flow in excess of pump capacity, or total flow if pumps are shut down, can overflow to a holding pond. Unacceptable product can be diverted to the same pond, as well as the contents of the clarifiers if they must be emptied for maintenance activities.


HOLDING POND
OVERFLOW
GRIT BASIN
PUMP STA
COLOR REMOVAL PROCESS
DRAINS
OUTLET BASIN
TO STREAM
BIO-OXIDATION SYSTEM
RAW WASTE

EMERGENCY FLOW SCHEME FOR EFFLUENT

Figure 12

SECTION V

EQUIPMENT PERFORMANCE AND CHANGES

Grit and Trash Removal

The grit conveying system, as delivered by the vendor, had several obvious operational shortcomings. Joints in the conveyor drag flight guides were vulnerable to firm hanging against the ends of conveyor drags. At the downstream extremity, these guides lacked the clearances necessitated by the lateral movement permitted by the chain tracking variations. Shape of these members encouraged jamming by sticks or large chips in the effluent.

During the early hours of trial running with low water levels (to facilitate visual operation), these deficiencies were identified. The parts were trimmed, shaped and bent to obtain suitable spacing and self-clearing characteristics.

The conveyor flights are bolted to conveyor chain lugs, and a number of loosened bolts were noted during early weeks of operation. All nuts were welded to the bolts to eliminate the obvious hazard involved.

After a few weeks of operation, conveyor drag flights exhibited noticeable wear at the points of contact with steel wear strips in the bottom of the grit basin. A welded overlay of hard alloy was applied, and wear has been reduced to an acceptable level. It appears that several years may elapse before these drag members must be replaced.

Subsequent performance has been largely trouble-free. On occasion, the over-night collection of settled material has exceeded considerably the capacity of the 6 cu. yd. refuse container. Most of this material would have been harmless to the process system. The most common type of material has been woody fragments which have been cut into granules (rather than defibered) by pulp refiners. Even some defibered pulp settles out under the conditions required for grit sedimentation. Good in-plant operation has usually been sufficient to keep the volume of settled material below troublesome levels.

The disc screen, after some initial problems in correcting a badly-formed seat for the rubber seal strip, has performed satisfactorily. The timing control was designed to provide a minimum 3-minute cycle for the screen and a longer period for the carry-off conveyor. It was found that, after the screen stopped, head loss across the screen would sometimes reach switch-on point before the carry-off conveyor had completed its cycle. Under such conditions, the timer system would not function until the manual stop-start was actuated. An electrical solution to the problem was found. However, it is felt the longer cycle for the conveyor is superflous, since the quantity of material on the conveyor would be insufficient (after three minutes of screen rotation) to cause difficulty, even under freezing conditions.

The disc screen has exhibited adequate hydraulic capacity, the one-half inch drilling has provided the needed system protection, and the quantity of debris collected has been so small as to present no problems.

Lift Station

The pumping station, as engineered, seems to be an adequate design. However, experience with the pumps has emphasized the importance of stringent and detailed specifications when low-bid acceptance is mandatory. In particular, the pumps have required excessive maintenance attention to shaft seals and to lower bearings. The seals have been extensively modified to counteract their tendency to loosen by turning in the direction of pump rotation.

Lime Slurry System

The lime slaking and feeding system started without difficulty and operated as expected. At first, the temperature rise of water from input to output did not seem to correspond to the expected relation between volumetric flow rate and weight of lime added. Analyses of slurry samples yielded results which corresponded to the temperature rise but not to the lime found in the receiving process; however, reproducibility of these data was poor. After several weeks of operation, it was discovered that the water flow meter scale was incorrect. The indicating scale was linear, while the instrument characteristic was logarithmic; also, full-scale flow was twice the maximum scale indication.

When the correct scale was used, agreement was found between temperature rise and hydrated lime content. For the reaction,

$$CaO + H_2O \longrightarrow Ca(OH)_2$$

heat release is 486 Btu per pound of CaO reacted. (20) Thus, slaking lime to produce a slurry containing 10% calcium hydroxide (7.6% CaO) will lead to a temperature rise of about 38°F. If twelve million gallons of effluent per day (8,333 g.p.m.) are to be treated with 1,000 mg/l (w/w) of CaO, the required slurry would be about 98 g.p.m. at a temperature rise of 38°F. It should be remembered that temperature differential reflects CaO actually slaked, not lime feed to the slaker.

Control of lime feed to the slaker has been affected by two types of mechanical problems. The first was choking of the feed conveyor with metal debris. The chief source of such material was broken and detached buckets from the lime bucket elevators, which were not adequate to the demands of the heavy, hot-lime service, due to specifications which were not sufficiently stringent in writing and/or interpretation. The other problem was performance of the variable-speed control on the feed screw. This control functioned by shifting effective diameters of pulleys in a belt-driven speed reducer. Failures occurred in the small motor which controls the drive ratio. Two solutions to the latter are: (1) use of a simple, manual-crank adjustment, or (2) a different or heavier-duty speed controller (both more expensive). Experience with this system suggests the cheaper, manual system. Only rarely have feed rates been changed more than once during an 8-hour operating shift.

During the early months of operation, little difficulty resulted from omission of any device for heating the slaker water supply. This might be partially due to the frequent availability of recovered process water at about 140^{o}F.; however, water at 80^{o}F. was also used part of the time. Later, periods occurred when slaking the incomplete and substantial amounts of lime were rejected by the classifier rate. It was usually confirmed by other evidence that lime reactivity was low on these occasions; commonly, the lime appeared "overburned", with hard, glazed pellets. Since such conditions will arise occasionally, it was found desirable to add a steam injection heater ahead of the water temperature measurement. By so controlling the heater that final slurry temperature is kept below boiling, the temperature rise can still be used to monitor lime feed. (Obviously, at higher temperatures, heat loss will increase.)

There had been concern that the long lime slurry delivery pipe would be plugged by solid lime deposits, since such experiences have been reported from other milk-of-lime pumping systems. However, the maintenance of good flow velocity has apparently produced the designed effect; no plugging of the main line has been observed. Deposition has occurred at three points in the slurry system: in the gravity overflow pipe from slaker to slurry tank, in the short suction piping to the slurry pumps, and at the point of slurry introduction into the mill effluent. The first two points represent low-velocity conditions (one pump is normally idle, with slurry in its suction pipe), and the third is a point at which calcium carbonate formation occurs. In all three, the periods between clean-outs has generally been three to five months. Piping has been arranged to facilitate quick punching out of the deposits. On one occasion, when there was a prolonged problem with low-reactivity lime, the classifier compartment of the slaker became plugged.

Color Clarifier

The "primary" or "color" clarifier mechanism functioned with virtually no difficulty, and with no outage for malfunction, throughout the 15 months service covered by this report. (It was used for 3 months as a conventional "primary" clarifier for mill effluent before start of the year of color removal demonstration.) This was true despite a much higher degree of sludge thickening and amount of sludge retention than contemplated in design. On one occasion, the torque switch, which was set low at about 1,000,000 ft.-lbs. torque, was tripped due to heavy sludge load. After about two hours, during which the torque switch was reset, the mechanism re-started and remained in service.

Distribution and flocculation in the input compartment ("center well") was satisfactory. At the close of the demonstration period, head loss measurements indicated some fouling of the inlet pipe, but the unit had not been shut down for inspection.

The clarifier shell, with discharge ports, collection launder, and overflow control dams, displayed satisfactory hydraulic capacity and control. Some fouling of discharge ports had required cleaning on two occasions. The material involved in the fouling was soft (easily removed) and appeared to be related to periods of inadequate lime dose, and to NSSC components and dried foam from the carbonation system. (This foam is discussed elsewhere.)

Carbonation Tank

When the carbonation tank was filled and the gas flow began, the four agitators were started, and the drive motors were found to be severely overloaded. The supplier of the agitators had warned that the motors would overload if the agitators were run in ungassed water, but should be within rating at normal gas flow. Trials were made with various gas flow rates to individual gas diffusers, and no condition was found which effected any considerable reduction in motor load. Each agitator had six blades which were bolted to a hub. By removing each alternate blade, the unit was left with three symmetrically located blades. With this change, the motor load was found to be within a few percent of full horsepower rating. Little difference was found between gassed and ungassed operation. Because of this last observation, it was possible to eliminate an electrical interlock between agitator motors and gas blower.

After the first three days of operation (interrupted at this point for other reasons), it was found that three of the four gas diffusers had broken loose at the welded "legs" which had secured them to the tank bottom. This occurrence is an indication of the violent turbulence which exists in the carbonator. Each diffuser was then welded to four new legs made of 3" x 3" heavy angle iron. The legs were full-perimeter welded to the tank bottom, and each was double-welded along a 3-inch contact with the diffuser. No further trouble was experienced with these anchors.

After about 50 - 60 days of actual operation, the interior of the carbonator was inspected. The agitator blades were covered with a hard scale of calcium carbonate which ranged up to more than one inch thickness. The diffusers were also coated with about one-half inch of scale along the lower edges, so that the serrations in the metal edges were largely rounded over. Scale on the agitator's shaft (steel) and gas supply pipe (stainless steel) was rather thin and showed evidence of intermittent flaking. The vertical tank wall had a deposit about 3/4" thick, which was soft enough to be scraped away with a thumb nail. Scale on the agitator blades and diffuser edges was hammered off in about twenty minutes. It was subsequently found desirable to repeat this removal about once per month. Other deposits in the tank proved self-limiting and have caused no serious trouble.

The tank was originally provided with two round vent stacks of 24" diameter. Soon after operation began, it became apparent that the exit gas velocity was sufficient to entrain any froth, with scant opportunity for water showers to break the bubbles. Substituting two 30" square stacks (approximately doubling the cross-sectional area) improved the situation, but at times foam emission was still a nuisance. The possibility of foam problems had been foreseen prior to construction, and there had been discussion of a mechanical foam breaker or a foam tower as potential devices to deal with foam. The expelled foam was observed to be heavy, wet and laden with calcium carbonate. A temporary expedient, which would give opportunity to examine behavior of the foam, was decided upon. A horizontal, 20" diameter pipe was run from the lower portion of one vent stack (total height, 8 feet) across the outer wall of the color clarifier. Water level was more than a foot below the top of the clarifier, providing capacity for a substantial volume of foam. It was found that the foam spread and decayed rapidly in contact with air. Any calcium carbonate or other solid matter was trapped and recovered by the clarifier. Although the partially-dried foam residue was unattractive in appearance, an adequate foam disposal was provided, and the usage has been continued.

After about four months service, a blade-mount stub on one agitator hub broke, apparently as a result of metal fatigue. A similar breakage was experienced the following month; this time, the break occurred at night, and before it was discovered, the resulting vibration wrecked the reduction gear housing. One additional hub failure was recorded during the first year of operation. During this period, there were two gear replacements due to broken gear teeth.

Toward the close of the year of operation covered by this report, some limitations in hydraulic flow capacity of the carbonator were observed. One observation indicated a change in the system; another resulted from increased volume handled, revealing a latent problem. In the first instance, it was found that scale formation in the discharge pipe was restricting outflow from the carbonator. The other case involved input flow: at higher throughput, an unexpectedly high liquid level was developed in the color clarifier outlet compartment. It appears that design was based upon liquid entry at the surface of water (specific gravity = 1.0), whereas in operation the liquid is gassed to a density of about 0.8 and rises some two feet higher. (Introducing flow at bottom of the carbonator would obviate the effect of gassing.)

Gas Blowers

The carbonation tank is supplied with lime kiln stack gas by a Roots-Connersville blower (or compressor -- output pressure is 5 - 10 psig) rated at 4,000 ACFM. Within the first several days of operation, this unit became the object of two serious concerns.

The first concern related to suitability of the equipment to the service demands. Bearings and seals required replacement at this time. Although examination indicated the failure was due to corrosion damage while the unit was idle between manufacture and start-up, several maintenance problems could be seen. Dust and scale accumulations (apparently calcium carbonate, chiefly) were already evident, and the seals seemed quite vulnerable to the effect of such solid material. Replacement of seals requires almost complete dismantling of the machine, and we have found it preferable (based on outage time) to move the entire unit to the plant machine shop.

Other seal failures have occurred, signalled by the appearance of water in the oil circulating system, with subsequent trip-out of the oil pressure switch. On one occasion, one of the head castings was cracked by pressure exerted by dust accumulations. There have also been problems re-starting the blower after it has been shut down for a short time; this was attributed to scale accumulations on the rotor lobes and housing. (Cooling reduced clearance below that required for free rotation.) Occasional injection of light fuel oil into the blower intake has reduced the scaling problem somewhat.

The second concern about the blower had to do with capacity. Even though effluent throughput during early operation was far below design flow rate, the gas supply was barely sufficient for carbonation at peak demand. This raised questions about the adequacy of design specifications.

To test the needs versus performance of the system, chemical analyses of gas and liquid flows were studied. During the early months of operation, it was found that the kiln gas was being diluted by air intrusion through a defective labyrinth seal between the kiln and the feed-end housing. The result was a CO_2 concentration of only 9 - 12% (dry basis). When the kiln seal was repaired, in the closing days of 1971, CO_2 concentration rose to a level of 16 - 17% (at minimum ratios of combustion air to the kiln). Exit gases from the carbonation varied in CO_2 concentration, depending upon pH of the liquid phase; however, the concentration was normally below 1.5% for pH above 10.0.

Carbon dioxide demand involves one factor which was not weighed adequately in design considerations. Much of the sodium content of raw waste is present as sodium-organic materials or sodium bicarbonate, which are involved in calcium precipitation reactions which produce sodium hydroxide, which in turn consumes carbon dioxide to reach normal carbonation pH; even more is needed for neutralization prior to bio-oxidation.

Review of the calculations of required kiln gas flow rate indicates three contributors to a low estimate. Soda alkalinity was one. Variations of calcium alkalinity above average estimates were not provided for. The third, and most serious, appears to be the assumption of kiln gas composition based upon fuel oil (carbon:hydrogen ratio of 1:2) instead of natural gas (carbon:hydrogen ratio of 1:4); no provision was made for variations from optimum sludge loading or fuel:air ratios.

Arrangements have been made to replace the blower with two units of the "water piston" type, having a combined capacity of 6,000 ACFM. The choice of type is based upon generally lower maintenance requirement and up-time as well as a lower vulnerability to dust content of the gas. In the latter regard, it should be noted that the normal dust content is very low because of the excellent venturi scrubber; however, momentary interruptions of scrubber function will entail substantial dust loadings.

One further note on the kiln gas supply system has to do with the pressure requirements for carbonation. The system was designed for a working pressure of 10 psig, with the expectation that normal back-pressure of the system would approach 80% of this value. Actual performance has involved full-flow pressure of about 5 psig. The cloud of gas bubbles above the gas diffusers, coupled with the flow induction of the turbine agitators, appear to result in discharge conditions more favorable than the calculated static pressure of the carbonation vessel.

Carbonation Clarifier

The second clarifier has functioned with no mechanical difficulties and no maintenance attention except routine lubrication. It had been expected that considerable torque would be developed if sludge withdrawal were not so controlled as to avoid large accumulatinns of sludge. It was assumed that the calcium carbonate produced in the carbonation step would behave similarly to that from kraft mill recausticizing; causticizing "mud" will settle to a dense cake which (at consistencies around 50% solids) can immobilize clarifier mechanisms. On the other hand, the sludge collected in the carbonation clarifier has not readily thickened beyond 25% consistency, and has poured from a sample container after sitting overnight. As a result of the sludge properties, torque developed by the clarifier drive mechanism has never exceeded 15% of the design rating.

Clarification performance of the unit has been less uniformly satisfactory. Losses of calcium carbonate in the clarifier effluent have ranged from 2% to 70%. The highest values have been associated with the presence of organic compounds high in ratio of NSSC-to-kraft origin, or with inadequate lime treatment. However, not all poor performance has been readily explainable in this manner. Efforts to improve retention by sludge recirculation -- either into the carbonator or into the clarifier influent -- have not yielded significant results. Flocculation tests with polyelectrolytes by "jar test" techniques, using samples from the clarifier center well, have not provided a solution to the problem. Visual observation of these tests does not indicate that the problem would have been eliminated by a moderate reduction in design rise rate.

Some scaling of the input pipe to the clarifier has been noted. It is believed this deposition is due largely to the frequently marginal completeness of carbonation, due to inadequate carbon dioxide supply.

Outlet Basin

The outlet basin receives the output flow from the carbonation clarifier, and the connecting pipe has caused some minor annoyance which suggests an improvement in piping design. Liquid level in the outlet basin is more than ten feet below that at the clarifier outlet; from the clarifier, the pipe extends several feet horizontally and then the effluent passes through a standard pipe elbow bend into a vertical pipe to another ell which directs flow into the basin below normal water level. Such a pipe configuration entrains large amounts of air and causes considerable splashing and may produce foam.

A chief function of the basin is to provide post-carbonation of effluent to reduce pH to a level compatible with subsequent bio-oxidation or in-plant re-use. The perforated pipe gas diffuser functioned well for a few weeks while enough CO_2 was provided for adequate carbonation plus pH adjustment. When carbonation became marginal, scaling of holes in the pipe began. It is felt the diffuser will be generally satisfactory when the gas supply problem is resolved.

Sludge Pumps

The pumps originally purchased to handle color clarifier underflow sludge were Morris Model 3JC14, a high-head sludge pump represented as capable of handling 1-1/4 inch spherical solids. When the clarifier (with the pumps) was put in service as a conventional primary clarifier, before the start of lime treatment of the installation of the disc screen, the pump intake quickly clogged with bark, wood slivers and pulp. A Morris 4 HS pump was borrowed to substitute for one of the original pumps and worked well in this interim, moderate-head service. The 4 HS pump as an impeller deeply recessed from the head (a style often described in the industry as a "trash pump"), providing high tolerance to rough solids, with a sacrifice of horsepower efficiency. The working head rating is 140', compared with 170' for the original pump.

When the entire system was completed and lime treatment begun, the remaining 3JC14 pump was again tried. When sludge consistency rose to a desirable range, recurrent choking at the pump suction was observed. The difficulty no longer seemed to be with coarse debris, but with rheological properties of the normal process sludge, which offered excessive flow resistance at the restricted inlet opening and led to cavitation at the impeller. On the other hand, the Model 4 HS pump worked quite satisfactorily, even though the sludge consistency averaged (contrary to earlier expectations) above 15% solids for monthly periods and exceeded 20% for several consecutive days. Often, the sludge required shaking to empty it from a 4-inch diameter cylindrical sample cup. (The pumping distance is over 650' through 6" pipe, terminating through a 4" magnetic flowmeter and 3" control valve. Intake is through about 90' of 8" pipe, with about 19' static suction head.)

The 1-1/2 JC14 pumps provided for the carbonation clarifier were also the source of some problems. When sludge accumulated in the clarifier, consistency would increase to about 20% solids; under this condition, the pumps would frequently fail to pump. Only by dilution could flow be maintained, although the concentrated sludge appeared very fluid; apparently the slurry is somewhat dilatant, and effective viscosity rises sharply at high shear rates at the restricted intake orifice of the pump. A 3JC14 unit removed from color sludge service was substituted for one of the pumps and has been used satisfactorily whenever sludge consistency was high.

The rubber-lined Galigher pumps which deliver color sludge to the centrifuge have worked reasonably well. The pumping system, which includes flow-control instrumentation with V-ball control valve, has been subject to flow-rate oscillations which are not desirable in a centrifuge operation. It is uncertain whether pump or valve is chiefly responsible; it has been suggested that a positive, progressing-cavity pump might provide better performance. However, there have not been enough difficulties to justify a change.

Carbonation Sludge Disposition

When lime treatment was begun and initial sludge pump problems resolved, some carbonation sludge was pumped to the recausticizing mud tank. There was no clear evidence that the new sludge altered filtration characteristics of the "mud" (nor certainty that it did not). Due to impurity load already in the system, in combination with causticizing equipment difficulties, the precoat filter was already unable to handle the existing load. The new liquid volume simply added to system losses.

To relieve this situation, at least temporarily, the carbonation sludge was diverted to the color clarifier input pipe. As expected, there was no noticeable change in clarification or color reduction. The change did involve an increase by 50 - 100% in the insolubles mass imposed upon the centrifuge, and the sludge thickened to a greater density. Torque loading on the color clarifier rakes was not markedly increased, and the centrifuge was able to tolerate the total sludge load. Since the arrangement proved workable and filter limitations continued, the change has remained in use.

Color Sludge Storage

The agitator mechanism of the color sludge storage tank was soon found to be much less effective in the sludge suspension than an identical unit handling recausticizing mud. When the tank was more than half full, a watery surface layer appeared, and when the tank was held 75% filled, with no addition or withdrawal, the torque increased until the drive mechanism overloaded. Even at low levels (30 - 40% full), solids caked along the lower walls of the tank, and performance of the withdrawal pumping system (to centrifuge) gave evidence of dense masses occasionally breaking loose.

After a consultation with mixing specialists, a supplementary, side-entering propeller mixer was installed at a level which did not interfere with the radial arms of the original agitator. The new mixer fully loaded a 20 HP motor. Located almost directly above the outlet pipe, the mixer has largely eliminated trouble from sludge lumps (although some deposits form on other areas of the tank periphery), and the tank can be filled without agitator overload or appreciable stratification of sludge.

Centrifuge

At start of operations, dewatering effectiveness of the Sharples solid-bowl centrifuge surpassed expectations. Since pilot-plant work indicated that every effort would be required to attain 25 - 28% consistency in the discharge cake, a low pond setting was employed. At some sacrifice of liquid retention time (expecting a consequent loss of centrate clarity), this setting provides a maximum drainage time for cake as it emerges from the liquid and is transported (by conveyor scroll action) up the "beach" comprising the conical discharge section of the centrifuge bowl. The first days of operation yielded cake at more than 40% solids.

After less than three days of operation, a centrifuge universal joint broke. Two universal joints connect the eddy-current brake with the differential spindle of the conveyor planetary gear box. (By controlled restraint of spindle rotation, the eddy-current brake regulates speed differential between the centrifuge bowl and the internal screw conveyor.) With failure of the universal joint, conveyor action ceased. The failure was not immediately detected, so sludge feed continued, with water being expelled while sludge solids accumulated within the bowl. Because of operator inexperience in detecting and diagnosing this type of occurrence, and failure to initiate prompt flushing and cleaning action, the conveyor became so tightly jammed that a screw flight was bent before removal was finally effected. During the next few weeks of intermittent system operation, two more universal joints failed, and there were a number of centrifuge trip-outs because of transient peaks in conveyor torque. By means of a drive pulley change, the centrifuge was slowed from 2,450 to 2,300 RPM; plate dams were changed to increasd "pond" level within the bowl; and a 5-second time delay was introduced into the high-torque trip-out switch wiring. With these changes, together with improved operator skills, satisfactory operation was soon established and continued for over two months before another universal joint failure.

During the following five months, universal joint failures gradually grew in frequency. It was noted that the difficulties were related to periods of high pulp fiber content in the sludge. Representatives of the centrifuge manufacturer, who worked closely with project personnel, had analyzed the problem as a "chattering" effect, involving rapid variations in conveyor torque, which result in metal fatigue. This conclusion was dramatically confirmed when a direct torque-arm restraint eliminating the cushioning effect of the eddy-current brake, was temporarily installed on the spindle shaft. Violent (and frighteningly loud) vibration quickly tripped the safety release.

The manufacturer's engineers now concluded that consolidation of the cake in the centrifuge bowl was leading to a situation in which large areas were alternately seizing and releasing as a unit, instead of rolling up locally along the edges of the conveyor scroll. They arranged to exchange the original conveyor for a new one with an altered scroll configuration. The effect of the change was a dramatic improvement. Torque indication dropped substantially and improved in uniformity. During the following three months, which concluded the demonstration period here reported, there were no further mechanical failures attributed to this problem.

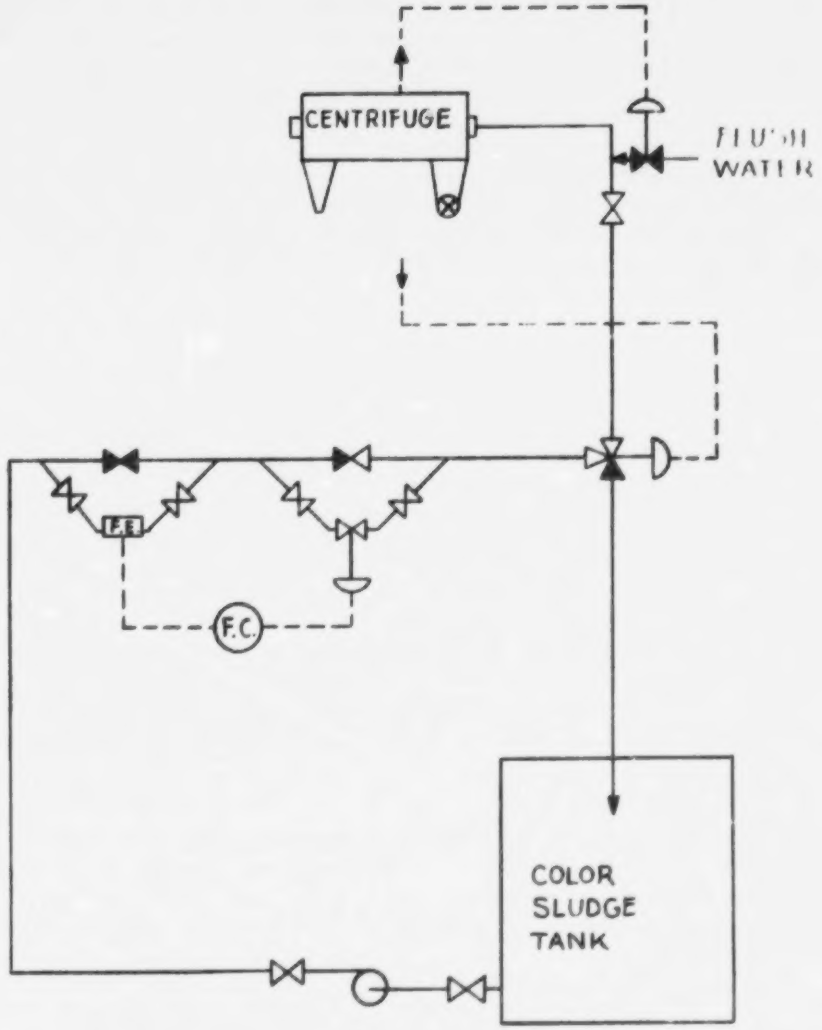
The original feed tube design provided with the centrifuge was a double-wall stainless steel unit which would permit use of flocculating additives (such use was never found necessary). There were several occasions of brittle fracture of these tubes. Substitution of a simplified tube design of mild steel seems to have avoided further failures.

The numerous protective circuits provided for the centrifuge (previously listed) have been the source of frequent interruptions of centrifuge operation. Some of the interruptions have been legitimate warnings of hazards, such as an unreliable cooling water supply pressure to serve the eddy-current brake. The high-torque trip was found to be responding to transient peaks for which its response was not helpful; a time delay provided a better control. However, there have been many spurious alarm signals. Trouble-shooting efforts have not clarified all of the problems; it appears that much can be attributed to a sophisticated and sensitive system operated in a humid and dirty environment. Relocation and protection of components is planned.

One control need which was not provided was a protective response in case conveyor action is interrupted, either by brake failure or by mechanical failure in the back-drive train (such as a universal joint). A planned addition is a relay to discontinue sludge feed whenever torque approaches zero.

One of the most commonly activated protective systems has been one to respond to rising torque load to interrupt sludge feed and inject flush water to prevent reaching the trip-out point. Since sludge feed rate is automatically controlled, an over-riding flow interruption distorts control action so that, when flow resumes, a brief period of cyclic over-response occurs. This has often triggered a repetition of the high-torque condition. To replace the use of a blocking valve to stop flow, a new piping arrangement has been designed to switch flow to a recirculation line; thus constant flow rate is maintained through the control loop. The new flow system is shown in Figure 13.

Centrifuge performance includes, in addition to cake dewatering, the clarity of the centrate. Since centrate is returned to the raw waste input, solids are not lost. However, if large proportions of the most troublesome solids are discharged with the centrate, they may be repeatedly recirculated until they accumulate to a quantity which cannot be


CENTRIFUGE
FLUSH WATER
F.E.
F.C.
COLOR SLUDGE TANK

CENTRIFUGE FLOW MODIFICATION

Figure 13

contained. It was therefore a matter of some concern when it was first observed that centrate consistency was sometimes more than 25% of the feed sludge consistency. At best, it was feared that polyelectrolyte flocculation aids would be required. However, no serious problem has arisen. Solids recovery by the centrifuge varied from 70% to 98% in a rather random time frame, with no evident cyclic pattern.

Successful sludge dewatering was one of the most critical engineering risks undertaken in embarking on the project. It now seems that the centrifuge application was a sound concept, and that persistent effort and cooperation by supplier and user have achieved a creditable working application.

Lime Kiln

Introduction of centrifuge sludge cake into the lime kiln has produced no evident problems except the obvious load limitations due to the moisture demands upon fuel and capacity. When the waste treatment sludge addition is started or stopped, there are no conspicuous differences in appearance of the product. When the kiln has been operated with this sludge alone for several hours, there have been no unusual observations. There have been a variety of problems in kiln operation, but all seem independent of whether waste treatment sludge was being processed. The original chain system installed in the kiln appears to have been more than adequate; it has now been shortened by burning back at the hot end. There has been some feeling that dust recirculation is higher with the effluent treatment sludge, but no data have yet shown clearly that this is so.

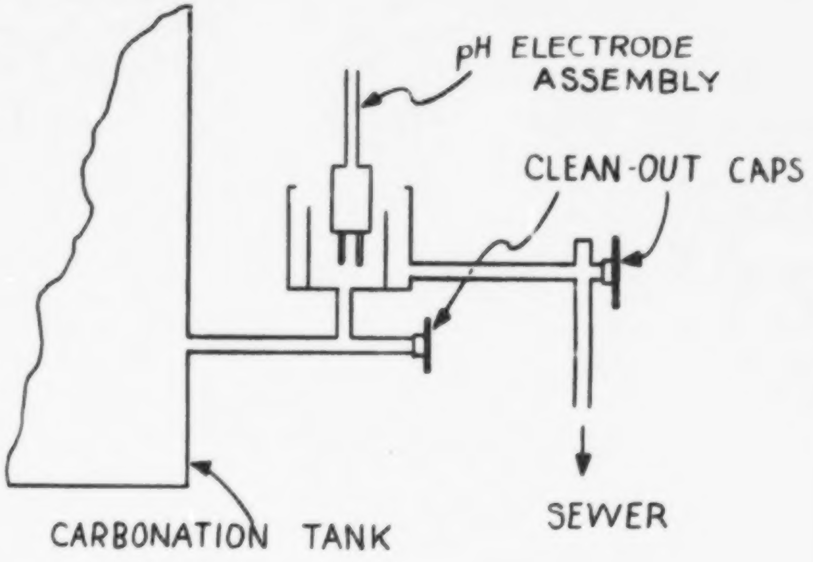
Instrumentation

In general, the instrument scheme has been found conceptually valid with the exception of some details of the centrifuge protective provisions and the disc screen control, both of which have been discussed in connection with the operation of the respective process equipment units.

One deficiency might be noted in the sizing of the control valve for sludge flow from the color clarifier to the color sludge storage tank (FV-014 on the instrument loop sheets). This 3" valve body with 2" trim has a very high pressure drop for a lengthy piping system which already presents a difficult pumping load, and there is no severe demand for narrow flow deviation. A standard 4" V-ball control valve would seem preferable and should avoid the frequent choke-ups which have required use of the 6" by-pass valve, which is too large for convenient manual control. (However, it has worked!)

One disappointing area of performance has been the pH instrumentation. Although some of the problems were due to calcium carbonate scaling of electrodes in the carbonation control system (see below), the pH measuring circuits themselves have performed rather poorly. The problems include standardization drift, non-linear response, and generally erratic behavior. In addition there has been frequent non-agreement between primary (transmitter) and remote indications.

As previously noted, calcium carbonate scaling -- of electrodes and of sample piping -- have been a problem in the carbonation pH system. The problem has been controlled by piping which can be easily, quickly and regularly rodded free of deposits. The electrode assembly, provided with flexible cable, can be readily lifted out of the chamber so that the electrodes can be cleaned with hydrochloric acid squirted from a squeeze-bottle. The sampling configuration is shown in Figure 14.


pH ELECTRODE ASSEMBLY
CLEAN-OUT CAPS
CARBONATION TANK
SEWER

FLOW SAMPLER FOR pH PROBE

Figure 14

SECTION VI

PROCESS PERFORMANCE

General Operating Considerations.

The Continental Can Company mill at Hodge, Louisiana produces unbleached kraft pulp (primarily pine) and a smaller amount of neutral sulfite semichemical (NSSC) hardwood pulp with "cross-recovery" whereby spent NSSC liquors are processed in the kraft recovery system. During the demonstration period covered by this report, paper machine output averaged about 750 tons per day. Typically, NSSC corrugating medium (79% NSSC, 21% kraft) was produced at about 300 tons per day during three periods each month, totaling 10 days. Thus, a month might include 8 days (rarely over 3 consecutive days) of 30 - 40% NSSC pulp consumption, 17 days (seldom over 7 consecutive days) of all-kraft production, and 5 days involving transition. Because of the varied product mix of the mill, daily production varied widely between 600 and 900 tons.

Effects of NSSC pulping upon effluent properties has been accentuated by the fact that, while kraft pulp washing performance at the Hodge mill was well above industry average, the NSSC washing facilities were much less effective. (Improvements were under construction, but not completed during this period.) As a result, NSSC contributions to effluent color were disproportionately large, compared to tonnage ratios. The varied production rates of the mill (noted above) also result in variations in the concentration of kraft components of the effluent.

Mill water use averaged around 12 million gallons daily (8,300 g.p.m.). Throughput of the color removal system was typically 6,000 - 7,000 g.p.m. Wastes from power plant, causticizing area, bag manufacturing and byproducts processing were excluded from plans for the system. Some waste water from the evaporator and recovery furnace area was not treated because needed sewer changes had not been completed. Liquid contents of each of the clarifiers is about 1,600,000 gallons above the settled sludge, so that the theoretical total detention time of the two clarifiers was around eight hours. However, in these deep clarifiers, the characteristic liquid travel is far from "plug" flow, and this fact will have sybstantial effects on concentration/time gradients through the system.

After some early attempts to establish a program of sampling to follow time lag through the system, a uniform daily sampling procedure (based on convenience) was adopted. Raw waste, and the overflow effluents from the two clarifiers were sampled by three continuous samplers from which 24-hour composites were gathered at the start of each working day. The raw waste was sampled by a 75ml dipper which was energized at flow-proportional intervals; this type sampler was chosen to assure true sampling of fiber content and to minimize interference of trash. The clarifier outputs were sampled with diaphragm-type metering pumps. "Grab" samples of sludges, lime and lime slurry were gathered at the same time, as were process instrument data.

The operational project, as just outlined, offers some impressive advantages, and also presents some limitations for conducting a technical evaluation study. Among the advantages:

1. The plant's operational and waste disposal programs were firmly committed to the project, assuring motivation for successful operation.
2. The day-to-day variability of the effluent tested the capabilities of the process under a very wide variety of conditions.
3. Unpredictability of some process demands placed many problems in the hands of operating crews revealing adaptability to "real world" resources.
4. The scale of the operation and the length of the demonstration period eliminated many of the questions commonly inherent in experimental operations.

Among the limitations encountered:

1. It was not possible to develop fixed experimental programs under constant conditions to afford ideal statistical bases for evaluation. Indeed, the variability of conditions necessitates much selectivity in data handling to determine some of the desired relationships. The time lag and mixing effects (for example) necessitate consideration of the previous day's parameters in evaluating an item of output quality data.
2. Capacity of equipment, and the necessity of handling all mill output, sometimes limit the exploration of significant ranges of key variables.
3. Because of the size and cost of some equipment components, system alterations require much time. Even with the most expeditious handling, procurement of some large items may require several months.
4. Some transient input parameters cannot be evaluated because the response peaks are so greatly attenuated by in-process mixing.

Start-Up and Operation.

Full-scale lime treatment was begun on August 18, 1971 and continued in promising fashion for three days before it was interrupted for more than a week by a mechanical failure (which has been discussed elsewhere). Reasonably continuous operation was established in very early October. Except for 10 days for holidays and repairs and alteration work in December, and 7 days lost in January because of lime kiln problems, general system up-time has been above 93%. The great majority of the lost time has been due to problems with lime kiln, lime conveyors and (to a lesser degree) kiln gas compressor.

During operation of the system, lime feed to the slaker has been maintained almost 93% of the time. The lime interruptions involved in this calculation ranged from several minutes to a few hours, and they were largely due to conveyor problems and to clogging at the outlet of the lime supply bin. The record by months is shown in Table 19.

During all of this operating period a comprehensive monitoring program was maintained. A sample of the daily data sheet is pictured in Figure 15. Monthly tabulations of raw data are presented in Appendix D. The test methods are found in Appendix C. A preliminary summary of the early months of operation has been published previously. (21)

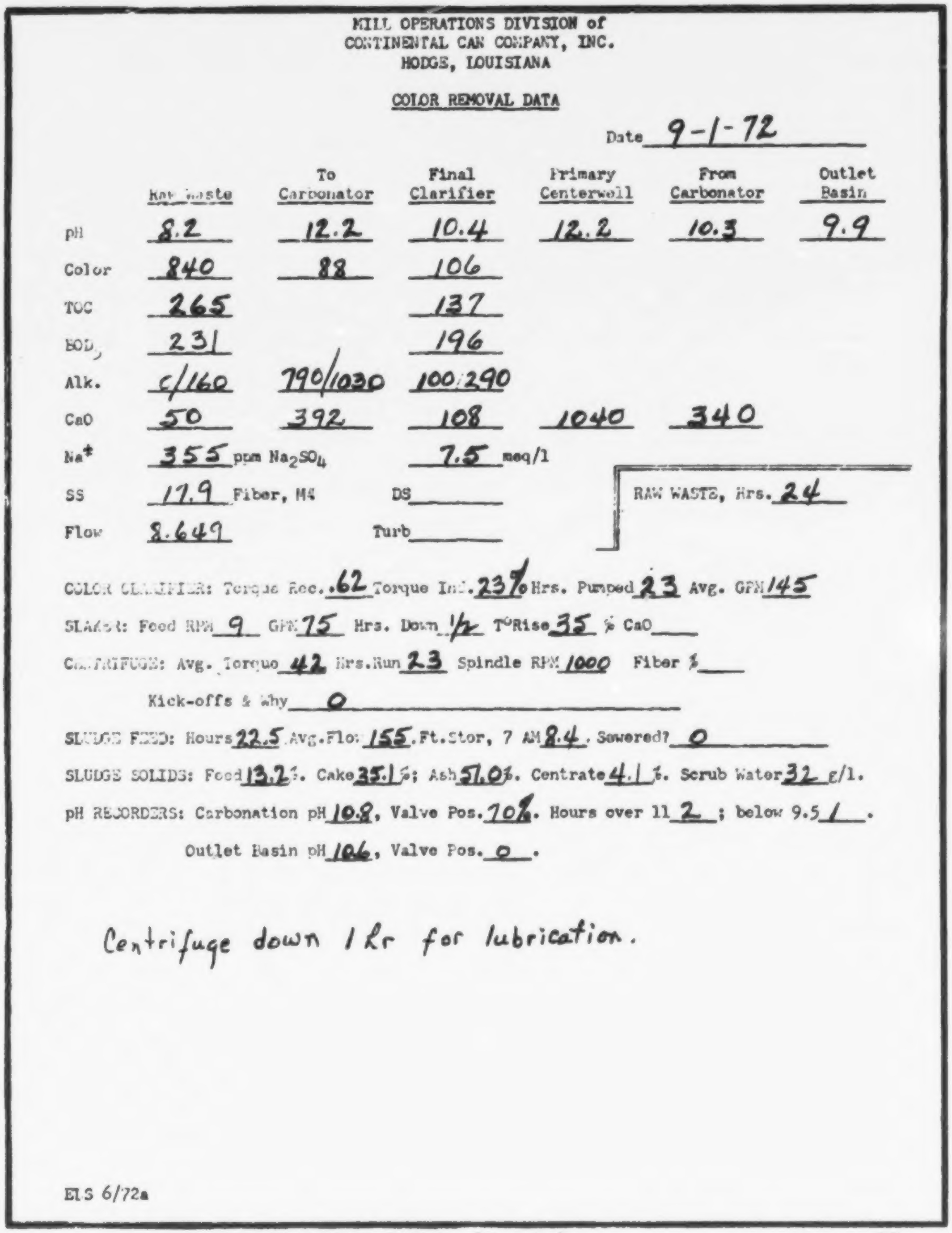
TABLE 19

TREATMENT SYSTEM UP-TIME

Month	Flow Up-Time, %	Lime Feed, % of Flow Up-Time
October 1971 (29 days)	100	91
November	100	96
December (21 days)	100	96
January 1972 (24 days)	86	83
February	99	95
March	85	88
April	91	93
May	82	97
June	95	93
July	91	96
August	95	91

Effluent Effects.

Reduction of effluent color, as measured in daily composite samples, has varied widely, ranging up to above 95% reduction. Although some of the occasions of poor color removal have not been associated with any recognized process aberration, most of the data can be explained by identifiable factors. The two most conspicuous factors have been inadequate lime addition and the presence of a high ratio of color derived from NSSC process wastes. There have been a few occasions when color sludge appeared in the overflow of the color clarifier and was



MILL OPERATIONS DIVISION of
CONTINENTAL CAN COMPANY, INC.
HODGE, LOUISIANA

COLOR REMOVAL DATA

Date 9-1-72

	Raw Waste	To Carbonator	Final Clarifier	Primary Centerwell	From Carbonator	Outlet Basin
pH	8.2	12.2	10.4	12.2	10.3	9.9
Color	840	88	106			
TOC	265		137			
BOD_5	231		196			
Alk.	c/160	790/1030	100/290			
CaO	50	392	108	1040	340	
Na^+	355 ppm Na_2SO_4		7.5 meq/l			
SS	17.9 Fiber, M4	DS				
Flow	8.649	Turb				

RAW WASTE, Hrs. 24

COLOR CLARIFIER: Torque Rec. .62 Torque Ind. 23% Hrs. Pumped 23 Avg. GPM 145

SLAKER: Feed RPM 9 GPM 75 Hrs. Down 1/2 T°Rise 35 % CaO ___

CENTRIFUGE: Avg. Torque 42 Hrs.Run 23 Spindle RPM 1000 Fiber % ___

Kick-offs & Why 0

SLUDGE FEED: Hours 22.5 Avg.Flow 155. Ft.Stor, 7 AM 8.4. Sewered? 0

SLUDGE SOLIDS: Feed 13.2%. Cake 35.1%; Ash 51.0%. Centrate 4.1%. Scrub Water 32 g/l.

pH RECORDERS: Carbonation pH 10.8, Valve Pos. 70%. Hours over 11 2; below 9.5 1.

Outlet Basin pH 10.6, Valve Pos. 0.

Centrifuge down 1 hr for lubrication.

ELS 6/72a

DAILY DATA WORK SHEET (Sample) Figure 15 partially re-dissolved. There appear to be a larger number of days on which lime feed was interrupted one or more times and, although the average lime feed was adequate, some of the color clarifier output during the day had received lower dosages.

Average color reduction for all days of system operation during the last eleven months of the demonstration period was 70%. If only the days of 100% kraft production (i.e., when no NSSC effluent was involved) are considered, the average color reduction was 80%.

Direct measurement of lime addition to the system was not precise enough for proper evaluation, so the selected indicator of lime treatment has been the soluble calcium content of the liquid overflowing the color clarifier. Indeed, theoretical considerations support the conclusion that this parameter is the prime determinant of color precipitation. Upon examining the data representing 40 days on which color reduction was above 88%, it was observed that, in virtually every case, the soluble calcium at the color clarifier had been 400 mg/l or more. Separate tabulations of data for kraft effluent and effluent including NSSC wastes were plotted to show the relation between percent color reduction and calcium ion concentration. A most-probable curve was then drawn for each set. These curves are shown in Figure 16 and 17.

Throughout the demonstration period, it has been noted that the average color of water samples taken at the color clarifier outlet has been lower than that of samples taken at the carbonation clarifier outlet. For the total period, this difference has amounted to an increase of about 25% in color after the first clarifier. The effect on this color difference on calculated color removal efficiency, for both all-kraft and NSSC-containing effluent is shown in the monthly averages listed in Table 20.

TABLE 20

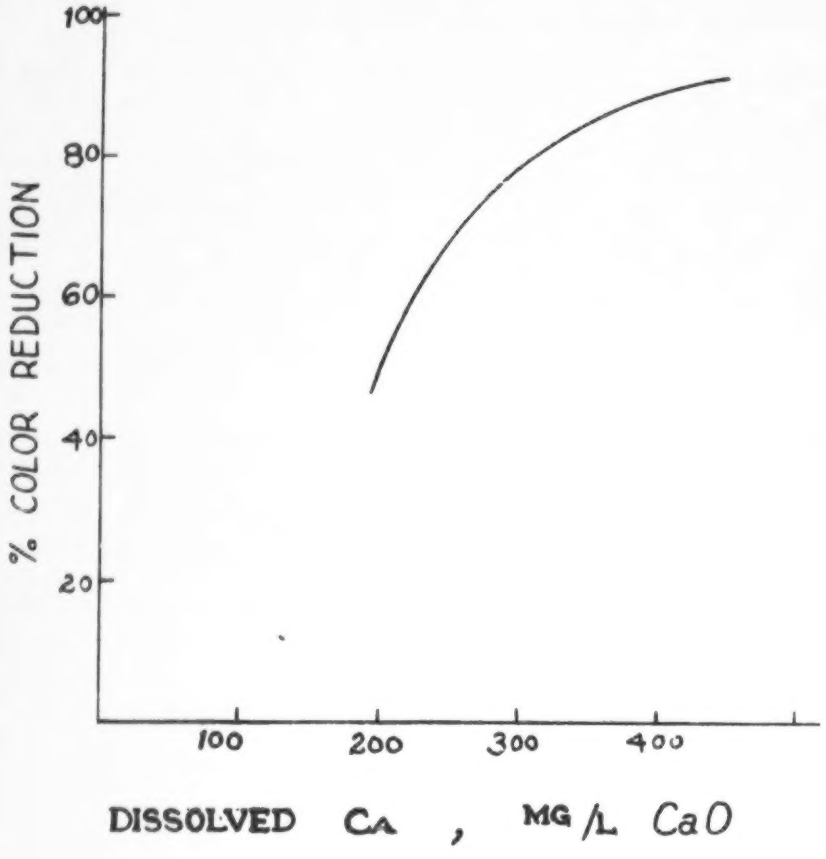
COLOR REDUCTION BEFORE AND AFTER LIME RECOVERY

	All-Kraft		With NSSC	
Month	Before[a]	After[b]	Before[a]	After[b]
October	83.3%	79.5%	70.0%	64.8%
November	85.1	82.1	61.7	58.6
December	85.7	83.1	64.3	57.9
January	78.0	73.5	67.7	64.1
February	89.6	82.1	69.2	64.0
March	84.6	76.6	73.6	67.7
April	87.9	79.4	78.6	70.7
May	82.0	79.2	73.6	63.2
June	88.6	85.9	71.2	65.5
July	86.1	84.0	75.8	68.1
August	83.4	74.3	71.3	66.8

[a] Sampled at outlet of color clarifier.

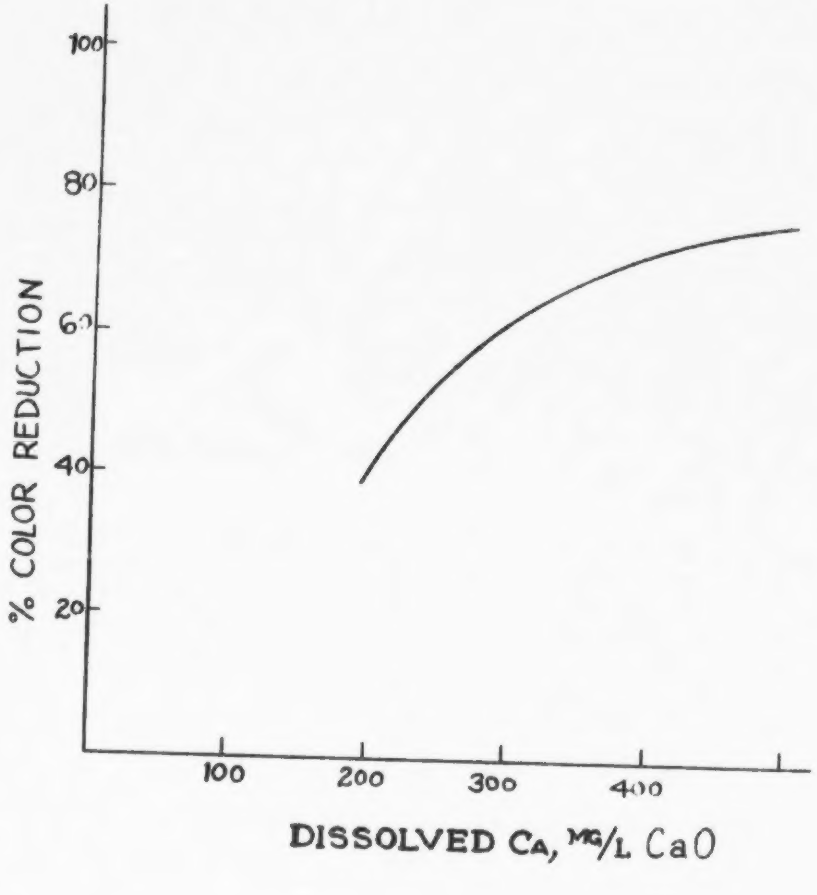
[b] Sampled at outlet of carbonation clarifier.

Percentage based on raw waste color.



SOLUTION CaO versus COLOR REMOVAL (KRAFT)

Figure 16



SOLUTION CaO versus COLOR REMOVAL (WITH NSSC)

Figure 17

There has been much speculation about the cause of this color increase.

Possibilities suggested have included:

1. Difference in chemistry of pH adjustment.
2. Sample storage at different pH and calcium levels.
3. Effect of unknown component of kiln gas.

The first possibility has been ruled out, but the cause is still not known. Further work is planned.

The reduction in total organic carbon (TOC) content of the effluent has appeared subject to the same factors as color reduction. When percent reduction is plotted against soluble calcium, there is much more scatter of data points. The scatter is believed to reflect a variable ratio of lignin to saccharides in the effluent. The low molecular weight soluble saccharides are largely unaffected by lime treatment. Curves representing data plots for all-kraft and NSSC-containing effluents are shown in Figure 18 and Figure 19. The data suggest that increased soluble calcium concentrations beyond 400 mg/l CaO will effect a greater proportional reduction in TOC than in color. Average reduction in TOC for eleven months was 37%.

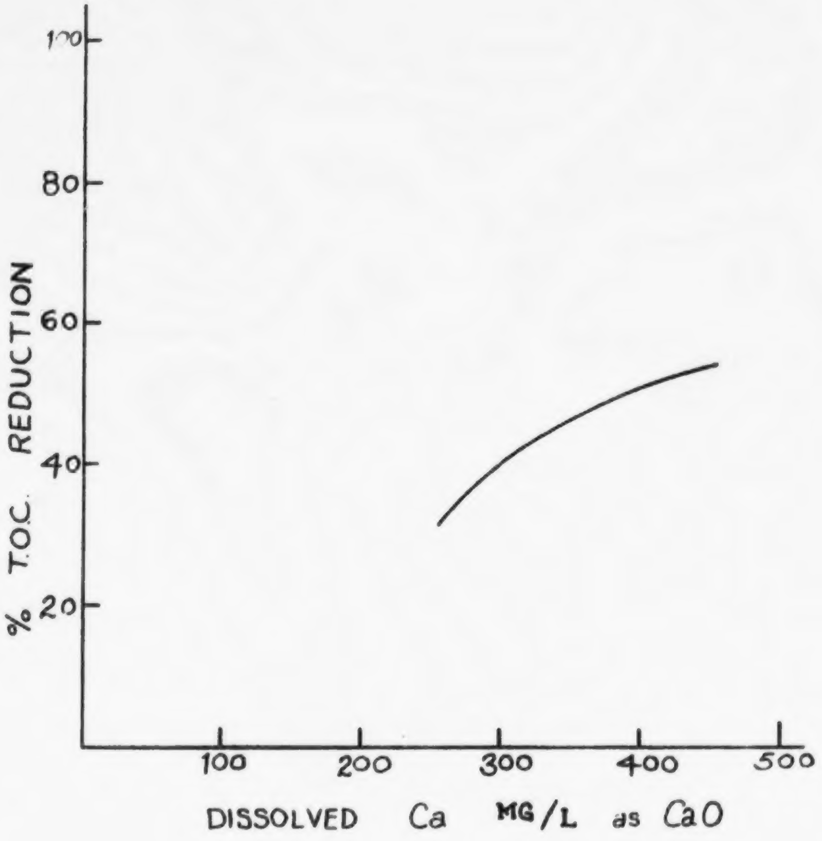
In correlating data on color and TOC, it was assumed at first that graphs relating the two would pass through zero on both co-ordinate axes. On this basis, it appeared the data yielded different slope lines for effluents with NSSC, as compared to all-kraft. However, closer examination indicated an intercept on the TOC axis at a value between 50 and 100 for the average of the data compiled in this project. Treating the data according to this hypothesis yielded an equation:

$$\text{TOC} = K + 0.19\ (\text{Color Units}),$$

which would apply to both groups. This positive, minimum value for TOC seems to reflect the fact that some lignin is precipitated by alum in kraft papermaking. Graphs reflecting project data for TOC and color are shown in Figure 20 and Figure 21.

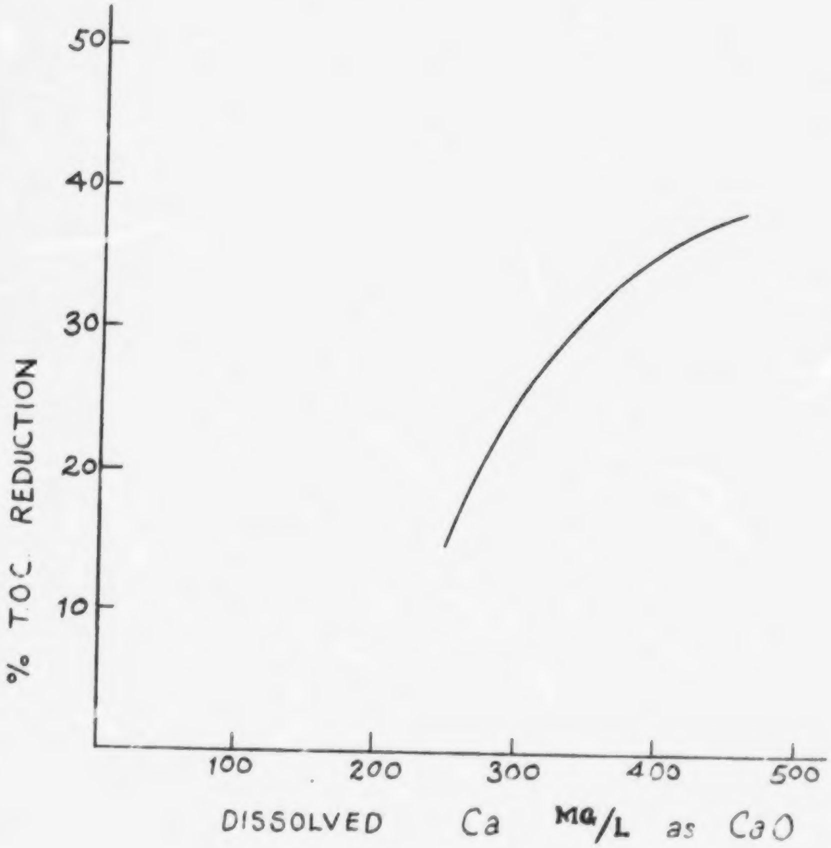
The effect of lime treatment on BOD_5 was one which did not bear out the promise of bench scale and pilot studies. A reduction of about 30 - 35% in BOD_5 had been expected, but no more than about 12% has been indicated by plant data. Meaningful information is not available for the full demonstration period, since sample collection during early months was found to be accompanied by enough bio-oxidation before testing to distort the results.

Fiber content of the raw waste ranged upward from 125 mg/l (24-hour average), and averaged about 300 mg/l. No levels were encountered which appeared to have any adverse effect on color removal.



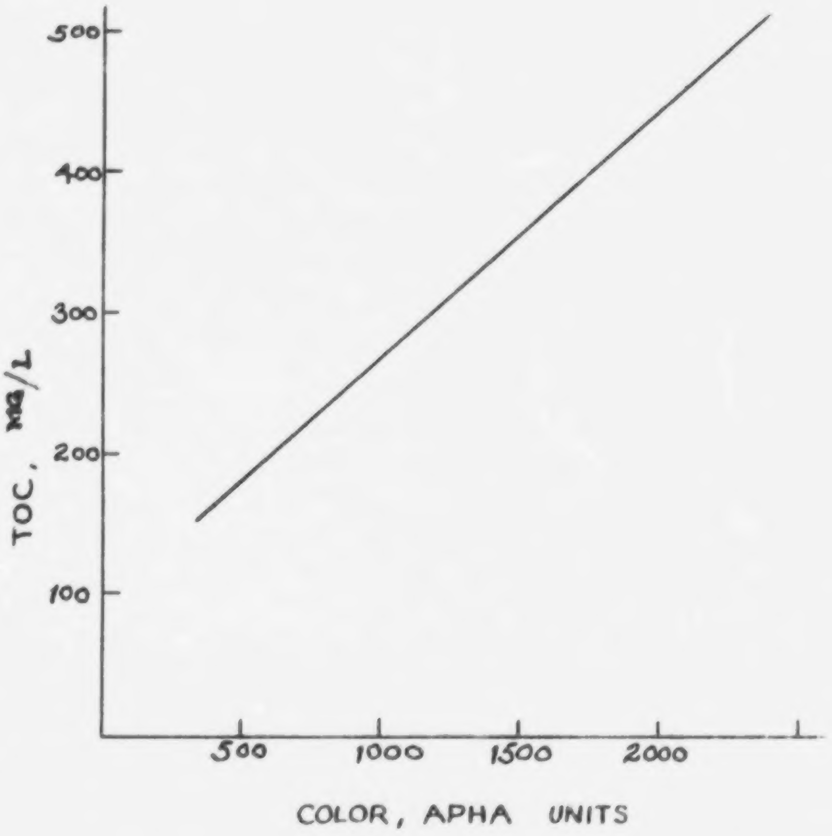
SOLUTION CaO versus TOC REDUCTION (KRAFT)

Figure 18



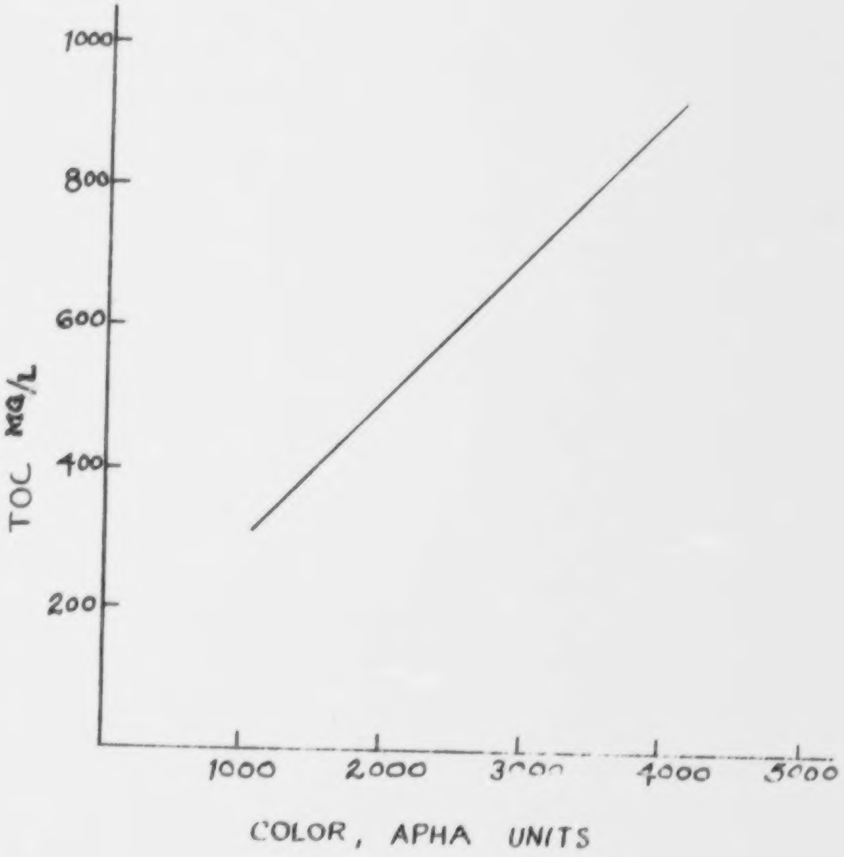
SOLUTION CaO versus TOC REDUCTION (WITH NSSC)

Figure 19



COLOR - TOC RELATIONSHIP (KRAFT)

Figure 20



COLOR - TOC RELATIONSHIP (WITH NSSC)

Figure 21

Clarification, Carbonation and Sludge Handling.

Clarity of color clarifier effluent has been excellent when adequate lime has been applied to kraft wastes. NSSC effluents sometimes have a slightly turbid appearance, although they are free of settleable solids. The only evidences of solids carry-over have been due to accumulation of excessive sludge volume in the clarifier. These occasions have resulted from sludges which did not thicken to the usual degree; the low rake torque which resulted was misinterpreted as indicating low sludge level, and sludge withdrawal rate was not adequate.

Sludge withdrawn from the color clarifier has averaged above 14% solids and has ranged from 9% to above 22%. The highest consistencies have been associated with high ratios of calcium carbonate, while the lowest consistencies have usually been related to high fiber ratio. Some of the variation has not been explained by such simple explanations, and there appear to be some more complex factors affecting sludge bulk and also influencing the rake torque developed at a given consistency.

Except for the mechanical limitations noted in the previous section, carbonation has proceeded in accordance with expectations. Sampling difficulties have limited the precision of estimates of gas absorption efficiency. However, it appears that the efficiency has been above 85% for pH values above 10.0. Efficiency will increase with pH. Automatic control of pH has required some careful instrument adjustment to compensate for the equilibration time of the system. However, control to within ± 0.2 pH units has been accomplished whenever proper instrument performance and steady kiln operation (uniform gas composition) were achieved.

Carbonation clarifier effluent clarity has been variable. Calcium losses in the clarifier overflow have ranged from 20 mg/l to over 200 mg/l as CaO; in the later months of operation, the average loss has been slightly above 100 mg/l as CaO, about twice the content of the raw waste. The most important adverse factors contributing to higher losses have been high quantities of NSSC components, and low lime dose. It appears that all-kraft effluent with adequate lime treatment would halve this loss. It is also believed that uniform pH control would assist settling of the calcium carbonate, although no pH has been established as best for this purpose.

Since underflow sludge from the carbonation clarifier is delivered to the color clarifier input, consistency is not critical, and a high pumping rate is normally maintained. Thus, there is usually no substantial sludge accumulation in the clarifier except in case of pump failure. As noted elsewhere, the limiting settled consistency has been about 25% solids, with surprisingly good fluidity. (See discussion in Section V.)

Centrifuge discharge cake has averaged about 33% solids over the full demonstration period. Since installation of the new design conveyor screw, the average consistency has been 35%, with a range of 30% to 40%. Centrate discharge has ranged from 1% to 7% solids, with an average consistency of 3.5%. At the average values of 14%, 35% and 3.5% for feed sludge, cake and centrate, respectively, the solids recovery of the centrifuge is 83.3%. It had been believed that, if average percent recovery fell so low, a retention aid would be required to avoid occasionally excessive accumulation of small-particle material. However, no additives have been used at any time.

Effects on Lime Burning.

The greatest uncertanties which had been felt about the effect of color sludge on lime burning were on two points: the possibility of forming rings or balls in the kiln, and the physical characteristics of the lime. In both cases, no effects have been noted. The kiln has formed balls, but the most serious were when no color sludge had been introduced. The product lime does not look different; there is no evident difference in fragility of pellets; reactivity in the slaker does not change when color sludge is added or omitted.

Although it seems certain that lime impurities are contributed by the color sludge, the amounts have been less than those involved in changes in the kraft mill system (specifically, changes in green liquor dregs handling). At most, the impurity level has risen to slightly over 10% (defined as the sum of acid insolubles and R_2O_3). There does seem to be some adverse affect on filtration of causticizing mud, but problems with green liquor quality have complicated evaluation.

The impact of color sludge on kiln capacity and fuel use has (at worst) not exceeded the calculated difference based on water content of the sludge. Precision of data on lime output has not been good enough to indicate whether the organic matter in the sludge makes a useful fuel contribution. However, observations at the time of fuel (natural gas) trip-outs have shown flame in kiln regions which should represent effective heat input.

Note:

A U. S. Patent (No. 3,639,206) has been issued, covering the process involved in this project. (22)

SECTION VII

ECONOMIC CONSIDERATIONS

Capital Cost.

Capital costs related to the color removal process may be divided into three categories: (a) process facilities used exclusively for carrying out the process, (b) cost of a proportion of lime-burning and related system for incineration and lime recovery, and (c) other costs, including land, sewer additions and alterations, holding reservoirs, access roads, etc. We shall deal here with only the first two.

The color removal process facilities are considered to include all effluent handling from inlet piping to the grit basin to the discharge piping of the outlet basin, including overflow and by-pass structures; and further to include lime supply and recovery, beginning at the lime feed conveyor to the color removal slaker and ending at the certrifuge discharge conveyor. An approximate subdivision of cost is as follows:

Lime Input	$ 35,602
Inlet Control; Grit & Trash System	121,319
Lift Station	55,170
Color Clarifier	274,335
Carbonation	95,994
Carbonation Clarifier	277,082
Outlet Basin	40,880
Sludge Storage & Dewatering	135,627
Instruments & Controls	115,015
Process Piping	256,278
Motor Controls	24,805
Electrical Wiring & Lighting	80,146
Pipe Bridge	51,547
Painting & Misc.	38,122
Spare Parts	59,393
Total Direct	1,661,315
Engineering, Plans & Specifications	87,825
Construction Supervision	12,421
Total Cost - Color Removal (Excluding Lime Kiln Cost)	$1,761,561

It is economically significant that the process performs the function of the conventional primary clarifier and provides complete and final sludge disposal by incineration. The color clarifier is actually smaller than the primary clarifier which would be needed for the same effluent. Thus, a new mill would require no other expenditures for these purposes, and an existing mill might adapt an existing clarifier to serve in the color removal process.

The lime kiln requirement to meet peak needs for effluent treatment amounts to at least 25% of total kraft mill lime supply. The cost of the 12' by

290' kiln totalled $2,667,000, including conveying and storage equipment. The effluent treatment share of cost would amount to $533,391. If additional calcining capacity necessitates an additional kiln, costs will be much greater.

Operating Costs.

During the demonstration period, the volume of waste treated was slightly below the acticipated flow rate, because of water use economies and because it was not possible to transfer some desired flows to the system. Cost factors have been developed for a flow rate of 9 million gallons per day and calculated for an average production of 750 tons of paper per day. The chief cost components are: lime loss, fuel use, electric power consumption and maintenance.

Lime loss has been calculated from the difference in calcium concentration between input and output water, and from sludge lost to sewer. During the last 7 months of the demonstration period, average lime loss was 85 tons per month. The data indicate that effluent losses have been high because of NSSC effects which have been previously discussed. There have been some sludge losses to sewer; with operational improvements, these losses should become rare. Costs have been calculated on the basis of purchased quicklime make-up at $22 per ton. Lime loss of 85 tons per month is equivalent to 7.5 pounds per ton of paper, or 630 pounds per million gallons. The costs are $0.0825 per ton, or $6.93 per million gallons.

Average operating electrical horsepower has been about 1,000, for which the cost of electricity is estimated at $100 per day. This amounts to $0.13 per ton of paper.

Fuel requirements attributable to color sludge processing have been equivalent to 17,500 million Btu per month. This calculates to 0.77 million Btu per ton, or 65 million Btu per million gallons of effluent. If a value of $0.48 per million Btu is assigned, the above requirements amount to $0.37 per ton of paper, or $31.12 per million gallons.

Maintenance costs are estimated at $50,000 per year; this is equivalent to $0.19 per ton, or $15.63 per million gallons.

Summarizing, operating cost (not including depreciation, taxes or insurance) under these particular conditions would be:

	Per Ton	Per Million Gallons
Lime Make-Up	$0.08	$ 6.93
Fuel	0.37	31.12
Electric Power	0.13	11.11
Maintenance	0.19	15.63
Total	$0.77	$64.79

Obviously, these costs will be much affected by mill tonnage and water use. To illustrate, we have assumed a mill expansion to 1,500 tons per day with water economies to yield a total water use of 15 million gallons daily, of which 13 million gallons requires lime treatment. Process balance and pulp washing are assumed adequate to avoid the NSSC effects on calcium carbonate recovery which applied to the previous case. Lime make-up needs are taken as 5,400 pounds per day. Treatment dosage and fiber loss are assumed to be in the same ratio to effluent volume as in the first case, resulting in a fuel demand of 25,280 million Btu per month (0.56 million Btu/ton). Electrical load is increased by use of an additional kiln gas blower and a second centrifuge which must operate at least half time, so that operating load is 1,350 HP. Total maintenance cost is essentially unchanged.

The resulting cost, based on the same unit prices as in the previous case are:

	Per Ton	Per Million Gallons
Lime Make-Up	$0.04	$ 4.57
Fuel	0.27	31.12
Electric Power	0.09	10.38
Maintenance	0.10	11.75
Total	$0.50	$57.82

It should be noted that the fuel pricing has a substantial effect on the total cost.

SECTION VIII

ACKNOWLEDGEMENTS

During a period of more than four years, many Continental people contributed to the successful completion of the project. Among them:

- Mr. S. Bruce Smart, Paper Group Executive Vice-President, headed the Continental management team, including Mr. E. A. Henry (Division General Manager) and Mr. J. G. Lee (Division Mgr. of Mfg.), which provided the leadership for a large commitment of corporate funds in a pioneering effort.

- Mr. C. C. Kunz, Plant Manager at Hodge, by insistent assertion of project priority, assured the needed team effort despite competing demands for time.

- Mr. W. Leroy Coker, Assistant Plant Manager of the Hodge mill, served as project manager during the major portion of the period.

- Mr. Edgar L. Spruill conceived the project, participated in the plant design and construction, and provided technical supervision of the operation and evaluation program.

- Mr. Fred Turner, research technician, performed much of the basic laboratory and pilot plant work which defined the project.

- Dr. John Schulz supervised process studies for the final plant design.

- Mr. Bobby Sammons supervised the departmental operating crews.

- Mr. Ross F. Miller co-ordinated the major efforts of plant design and construction.

The advice and support of Mr. Herbert Berger, of the National Council for Air and Stream Improvement, is appreciatively acknowledged.

The efforts of Mr. Dick Moll, Pennwalt Corporation, in helping establish satisfactory centrifuge dewatering are noted with thanks.

The encouragement and help of Mr. Robert A. Lafleur, Executive Secretary of the Louisiana Stream Control Commission, is gratefully acknowledged.

The support of the project by the Environmental Protection Agency, and the assistance provided by Mr. George Webster, Dr. James D. Gallup, Mr. Robert Hiller and the Grant Project Officers, Mr. George Putnicki and Dr. Richard Hill, are acknowledged with sincere thanks.

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SECTION X

GLOSSARY

ACFM - Actual cubic feet per minute; volume (including water vapor) at the existing pressure and temperature.

Black Liquor - Spent liquor from wood pulping by the kraft process, including dissolved organic matter and other products of cooking liquor reactions. (The organic content comprises about half the original dry weight of wood.)

BOD - Biochemical oxygen demand; BOD_5 refers to measurement of demand during a five-day period under specified conditions. (See Appendix C.)

Brown Liquor - Spent liquor from wood pulping by the NSSC process.

Carbonation - Treatment with carbon dioxide. In this paper, refers especially to neutralization of alkalinity and/or precipitation of calcium carbonate with carbon dioxide.

Caustic Extraction - A processing stage, in pulp bleaching practice, wherein lignin previously treated with chlorine or other oxidants is solubilized with alkali (usually sodium hydroxide) and separated from the pulp.

Causticizing - (Sometimes "recausticizing") - Conversion of sodium carbonate to sodium hydroxide by reaction with lime ($Na_2CO_3 + CaO + H_2O = 2\ NaOH + CaCO_3$); especially applied with reference to sodium carbonate recovered by incineration of kraft black liquor.

Clarifier- A device, usually continuous in operation, in which suspended matter is separated from a liquid by sedimentation, permitting withdrawal of a clear liquid a relatively concentrated slurry of the solid matter.

COD - Chemical oxygen demand, generally considered to represent the oxygen required to convert most organic substances to carbon dioxide, water and other fully oxidized compounds; commonly measured by dichromate consumption. (See Appendix C.)

Kraft - A widely-used wood pulping process employing sodium hydroxide with sodium sulfide to dissolve lignin. Also called "sulfate process," because sodium sulfate is used as a make-up chemical; carbon char in black liquor incineration reduces the sulfate to sulfide.

Mud (or "Lime Mud") - A common kraft pulping usage to designate the concentrated calcium carbonate slurry separated in the causticizing process. Used in this report to help distinguish from sludges or slurries generated by the waste water treatments described.

mg/l - Milligrams per liter; for dilute acqueous solutions, this value is approximately identical to parts per million by weight.

NSSC - Neutral sulfite semi-chemical, a pulping process employing sodium sulfite with (usually) sodium carbonate; hardwoods are commonly pulped at 70 - 80% yields for production of corrugating medium.

Recovery System (kraft pulping) - A process wherein kraft black liquor is concentrated by evaporation, burned in a furnace to recover heat value of organic substances (for steam generation) and chemical value of inorganic materials (which are ashed, dissolved and causticized to make fresh "cooking" liquor).

TOC - Total organic carbon, a measure of the carbon contained in the organic compounds in a waste water, for a given waste water source, this measurement may provide a useful index to chemical oxygen demand, or otherwise constitute a measure of pollution potential. This parameter is susceptible to a standard instrumental procedure. (See Appendix C.)

SECTION XI

APPENDICES

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APPENDIX A

EQUIPMENT LIST

Equipment No.	Name of Equipment	Mfr. & Model	Capacity, HP, etc.
0420	Grit Collector, with Chain Drive	Jeffrey Mfg. Co.	
0421	Motor	(Included)	3 HP @ 1800 RPM
0422	Speed Reducer	(Incl.) Falk 6K24	
0430	Disc Screen, Chain Drive & Reducer	Jeffrey Mfg. Co. (Steel exc. 31655 screen pl.)	14' Ø w/0.5" holes
0431	Motor	(Included)	1 HP @ 1800 RPM
0440	Conveyor, Screen Discharge w/drive	Jeffrey Mfg. Co.	
0441	Motor	(Included)	1 HP @ 1800 RPM
0010	Raw Waste Lift Station	Field Erected Sheet pile & concrete	18 Ø, 17' deep
0020	Raw Waste Lift Pump #1	Goulds Pump, Inc. VIM-14	6500 GPM @ 45'
0030	Raw Waste Lift Pump #2	Goulds Pump, Inc. VIM-14	9000 GPM @ 30'
0040	Raw Waste Lift Pump #3	Goulds Pump, Inc. VIM-14	Same as Pump #1 & #2
0021	Motor, Lift Pump #1	General Electric	100 HP @ 1200 RPM
0031	Motor, Lift Pump #2	General Electric	100 HP @ 1200 RPM
0041	Motor, Lift Pump #3	General Electric	100 HP @ 1200 RPM
0050	Color Clarifier/Thickener	Shell field erected concrete	135' Ø, 15' side-wall depth
0060	Clarifier Mechanism	Eimco Corp. CXXT Extra Heavy Duty	1,200,000 ft.-lbs. working torque
0061	Motor, Rake Drive, #1	General Electric	10 HP @ 1800 RPM
0062	Motor, Rake Drive, #2	General Electric	10 HP @ 1800 RPM

Equipment No.	Name of Equipment	Mfr. & Model	Capacity, HP, etc.
0070	Color Clarifier Sludge Pump #1	Morris Mch. Works 3JC14 (V-belt drive)	300 GPM @ 170' (500 GPM @ 40')
0080	Color Clarifier Sludge Pump #2	Morris Mch. Works 3JC14 (V-belt drive)	300 GPM @ 170' (500 GPM @ 40')
0071	Motor, Sludge Pump #1	General Electric	40 HP @ 1800 RPM
0081	Motor, Sludge Pump #2	General Electric	40 HP @ 1800 RPM
0100	Color Sludge Storage Tank	Rothschild Boiler & Tank Works	30 Ø x 20' Steel
0110	Tank Mechanism	Eimco Corp. Type BRM	
0111	Motor	General Electric	15 HP @ 1800 RPM
0120	Color Mud Pump #1	Galigher 2VRG200 2 x 2-1/2, belt-driven	185 GPM @ 80'
0130	Color Mud Pump #2	Galigher 2VRG200 2 x 2-1/2, belt-driven	185 GPM @ 80'
0121	Motor	General Electric	15 HP @ 1800 RPM
0131	Motor	General Electric	15 HP @ 1800 RPM
0140	Centrifuge	Pennwalt; Sharples P-5400	"x" 285 GPM
0141	Motor	General Electric (Spec. Winding)	200 HP @ 1800 RPM
0142	Eddy-Current Brake		40 HP
0150	Filtrate Hopper	Rothschild Boiler & Tank Works	
0160	Sludge Hopper	Rothschild Boiler & Tank Works	
0170	Centrate Head Tank	Rothschild Boiler & Tank Works	
0180	Lime Feeder with v.s. Reducer	Screw Conveyor Corp.	14" Ø x 16'
0181	Motor	(Incl. with reducer)	5 HP @ 1800 RPM

Equipment No.	Name of Equipment	Mfr. & Model	Capacity, HP, etc.
0200	Lime Slaker Conveyor with Reducer	Screw Conveyor Corp. Falk 323 EX II	15" Ø x 16'
0201	Motor	(Included above)	3 HP @ 1800 RPM
0210	Lime Slaker, with Rake Classifier	Dorr-Oliver No. 7	
0211	Agitator, Motor	General Electric	15 HP @ 1800 RPM
0212	Classifier Motor	General Electric	2 HP @ 1800 RPM
0220	Vent Stack	Rothschild Boiler & Tank Works	
0230	Grits Chute	Rothschild Boiler & Tank Works	
0240	Lime Slurry Tank	Rothschild Boiler & Tank Works	12' Ø x 10'
0250	Agitator, Slurry Tank	Cleveland Mixer Co. Heavy Duty Model AL-6	
0251	Motor	General Electric	7-1/2 HP @ 1800 RPM
0260	Lime Slurry Pump #1	Galigher Vac-Seal 2 x 2-1/2	180 GPM @ 55'
0270	Lime Slurry Pump #2	Galigher Vac-Seal 2 x 2-1/2	180 GPM @ 55'
0261	Motor	General Electric	10 HP @ 1200 RPM
0271	Motor	General Electric	10 HP @ 1200 RPM
0280	Carbonation Tank	Rothschild Boiler & Tank Works	30' Ø x 12'
0290	Agitator #1, Carbonation Tank	Chemineer, Inc. MDP-400-721	84 RPM
0300	Agitator #2, Carbonation Tank	Chemineer, Inc. MDP-400-721	84 RPM
0310	Agitator #3, Carbonation Tank	Chemineer, Inc. MDP-400-721	84 RPM
0400	Agitator #4, Carbonation Tank	Chemineer, Inc. MDP-400-721	84 RPM
0291	Motor, Agitator #1	General Electric 324 T	40 HP @ 1800 RPM

Equipment No.	Name of Equipment	Mfr. & Model	Capacity, HP, etc.
0301	Motor, Agitator #2	General Electric 324 T	40 HP @ 1800 RPM
0311	Motor, Agitator #3	General Electric 324 T	40 HP @ 1800 RPM
0401	Motor, Agitator #4	General Electric 324 T	40 HP @ 1800 RPM
0320	Compressor, Kiln Gas with Lube Oil System	Roots-Connersville 1421, RGVS	4000 ACFM
0321	Motor, Compressor	General Electric	250 HP @ 900 RPM
0322	Motor, Oil Pump	(with Compressor)	1 HP @ 1800 RPM
0330	Carbonation Clarifier	Tank, field-erected concrete	135' Ø 15' side-wall depth
0340	Clarifier Mechanism	Eimco Corp., CXXT Extra heavy duty	1,200,000 ft.-lbs. working torque
0341	Motor, Rake Drive, #1	General Electric	10 HP @ 1800 RPM
0342	Motor, Rake Drive, #2	General Electric	10 HP @ 1800 RPM
0350	Carbonation Sludge Pump #1	Morris Mch. Works 1-1/2 JCL4 (Belt drive)	100 GPM @ 120'
0360	Carbonation Sludge Pump #2	Morris Mch. Works 1-1/2 JCL4 (Belt drive)	100 GPM @ 120'
0351	Motor, Sludge Pump #1	General Electric	10 HP @ 1800 RPM
0361	Motor, Sludge Pump #2	General Electric	10 HP @ 1800 RPM
0370	Outlet Basin	Concrete, field-erected with gas diffuser piping	21' x 27' x 12' D.
0380	A/C Unit, Motor Control Room	Carrier Corp.	
0381	Motor, Compressor	(Included)	10 HP
0382	Motor, Fan	(Included)	1-1/2 HP
4000	Motor Controls	General Electric	
6000	Instruments & Process Controls	Foxboro Company	
6010	Instrument Panel, Control Room	Foxboro Company	

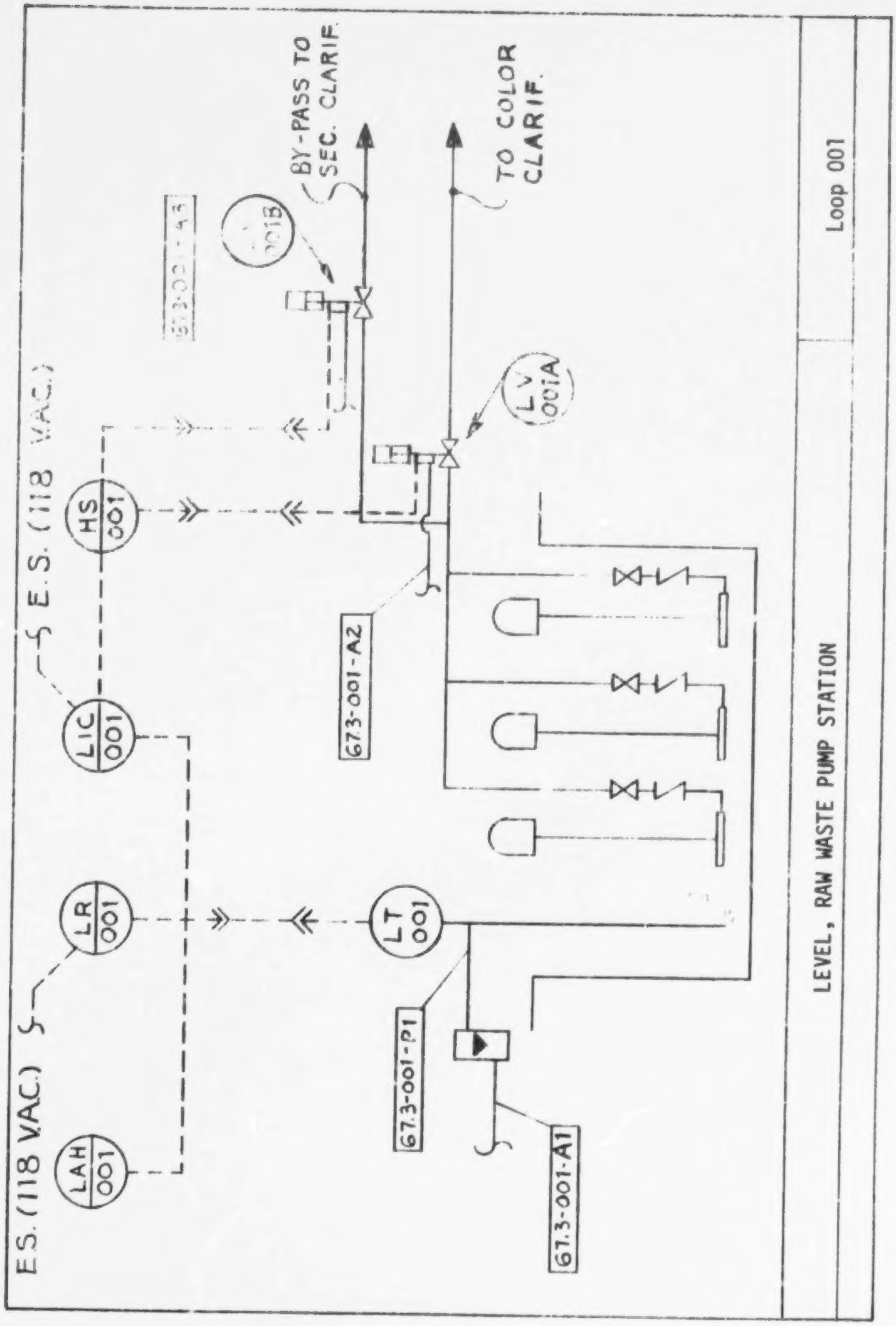
APPENDIX B

INSTRUMENT LOOPS

Loop 001	Level, Raw Waste Pump Station
LIC-001	Controller, 0 - 15 ft.; Proportional & Reset
LR-001	Recorder; 4 in. strip; 0 - 15 Scale
LT-001	Transmitter; 0 - 180 in. H_2O; output 10-5 MADC
LV-001A	Control Valve; 24" Butterfly; 316 SS Trim; Cyl. Op. w/ positioner
LV-001B	Control Valve, alternate flow - Sec. Clarifier; same as LV-001A
LAH-001	Annunciator - High Level
Loop 002	Flow, Raw Waste to Color Clarifier
FE-002	Orifice Plate, radius taps; 50" wc at 15,000 GPM
Loop 003	Speed Control, Lime Feeder
HS-003	Switch
SC-003	Speed Control
Loop 004	Speed, Lime Feeder
SI-004	Indicator, 0 - 30 RPM
ST-004	Transmitter
Loop 006	Dilution to Slaker
FI-006	Indicator, 0 - 200 GPM
FV-006	Control Valve, 2.5" CI Globe; w/positioner
FE-006	Orifice Plate, flange taps; 100" wc at 200 GPM
FT-006	Transmitter
FC-006	Controller
FAL-006	Annunciator - Low Flow
Loop 007	Temperature, Slaker Dilution
TE-007	Resistance Bulb, 0 - 300°F.
TR-007	Recorder, 4" strip; 0 - 300; 2 pen (with TR-008)
Loop 008	Temperature, Slaker
TE-008	Resistance Bulb, 0 - 300°F.
TR-008	Recorder, 4" strip; 0 - 300; 2 pen (with TR-007)
Loop 009	Lime Slurry Tank Level
LT-009	Transmitter, 0 - 120" water
LIC-009	Level Controller
LV-009	Control Valve, 2" CI Globe; diaphragm operator
FV-009	Flush Valve, Slurry Line; 2" DI Ball; Elec. 2-pos. operator
Loop 010	Color Clarifier Rake Torque
XT-010	Torque Transmitter

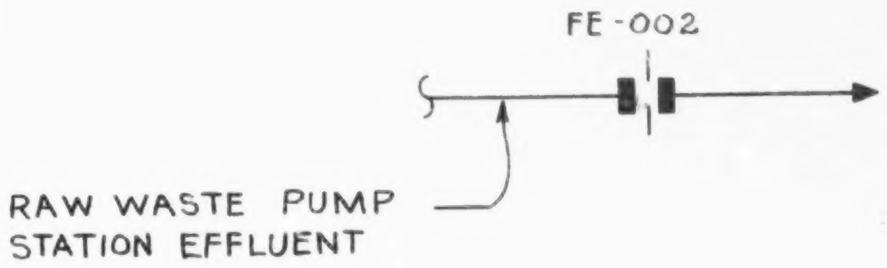
XR-010	Torque Recorder, 4" strip; scale 0 - 2 (with XR-011)
XS-010	Clarifier Rake Switch
XAH-010	Annunciator, High Torque
Loop 011	Carbonation Clarifier Rake Torque
XT-011	Torque Transmitter
XR-011	Torque Recorder, 4" strip; scale 0 - 2 (with XR-010)
XS-011	Clarifier Rake Switch
XAH-011	Annunciator, High Torque
Loop 012	Carbonation pH
AE-012	pH Element, Glass/calomel
AIT-012	Transmitter, pH, 0 - 14
AIC-012	Controller, with proportional and reset functions
AV-012	Control valve, 8", 316 SS Butterfly; diaphragm oper., w/ positioner
AR-012	Recorder, pH, 0 - 14; 4" strip (2 pen, with ZR-012)
ZR-012	Recorder, valve position; with AR-012
Loop 013	Outlet Basin pH
AE-013	pH Element, 014
AIT-013	Transmitter, pH, 0 - 14
AIC-013	Controller, with proportional and reset functions
AV-013	Control Valve, 3", 316 SS Butterfly; diaphragm oper., w/ positioner
AR-013	Recorder, pH, 0 - 14; 4" strip (2 pen, with ZR-013)
ZR-013	Recorder, Valve Position, with AR-013
Loop 014	Color Clarifier Sludge Flow
FE-014	Magnetic Flow Meter, 4", 304 SS Tube, Teflon Liner
FIC-014	Flow Controller, 0 - 300 GPM; Proportional & Reset Functions
FR-014	Recorder, 0 - 300, 4" strip; 3 pen, with FR-015 & 018
FT-014	Transmitter, 0 - 300 GPM; 10-50 MADC
FV-014	Control Valve, ball, 3" body, 2" SS Trim; Diaph. Op. w/ positioner
FV-014A	Control Valve, ball, 6", 316 SS Trim; cyl. operator
FAL-014	Annunciator, low flow
Loop 015	Carbonation Clarifier Sludge Flow
FE-015	Magnetic Flow Meter, 2", 304 SS Tube, Teflon Liner
FIC-015	Flow Controller, 0 - 150 GPM; proportional & reset
FR-015	Recorder, 0 - 150; 4" strip; 3 pen, with FR-014 & -018
FT-015	Transmitter, 0 - 150 GPM; 10-50 MADC
FV-015	Control Valve, 3" ball, 316 SS Trim; diaph. op.; w/ positioner
FV-015A	Control Valve, 4" ball, 316 SS Trim; cyl. operator
HIC-015A	Controller, hand loader, 0 - 100%
Loop 017	Color Sludge Storage Tank Level
LT-017	Transmitter, 0 - 240" water; 10-50 MADC
LR-017	Recorder, 0 - 20 ft.; 4" strip

LAH-017	Annunciator, high level
LAL-017	Annunciator, low level
Loop 018	Flow, Color Sludge to Centrifuge
FE-018	Magnetic Flow Meter, 4"; 304 SS tube, Teflon liner
FIC-018	Flow Controller, 0 - 200 GPM; proportional & reset
FR-018	Recorder, 0 - 200; 4" strip; 3 pen, with FR-014 & -015
FT-018	Transmitter, 0 - 200 GPM; 10-50 MADC
FV-018	Control Valve, 4" ball, 2" SS trim; diaph. op., w/ positioner
Loop 019	Annunciator
ANN-019	Annunciator, 4 wide x 3 high; horn & flasher
Loop 020	CO_2 Compressor Relief
PIC-020	Controller, 0 - 15; proportional & reset
PT-020	Transmitter, 0 - 15 psig; 10-50 MADC
PCV-020	Control Valve, 6" Butterfly, 316 SS; diaphragm oper., w/ positioner
Loop 021	Torque - Centrifuge
XR-021	Recorder, 0 - 100%, 4" strip
XT-021	Transmitter 0 - 50 mv to 10-50 MADC
XAH-021	Annunciator, high torque
Loop 023	Slaker Agitator Motor
QA-023	Annunciator, agitator stopped
Loop 026	Flush, Color Sludge Pump and Lines
FV-026	Control Valve, 1" ball, 316 SS Trim; diaphragm oper., w/ solenoid, 3 ways
FV-026A	Control Valve, 10" ball, 316 SS trim; cyl. oper., w/hand jack
HS-026	Switch
Loop 027	Flush, Centrifuge
FV-027	Control Valve, 3" globe, 316 SS trim; diaph. oper., w/ solenoid
HS-027	Switch
Loop 028	Carbonation Sludge Recirculation
FV-028	Control Valve, 3" ball, 316 SS Trim; cyl. oper.; w/ positioner
HIC-028	Controller, hand loader, 0 - 100%


E.S. (118 V.A.C.)
LAH 001
LR 001
LIC 001
E.S. (118 V.A.C.)
HS 001
LT 001
67.3-001-P1
67.3-001-A1
67.3-001-A2
LV 001A
001B
67.3-001-A3
BY-PASS TO SEC. CLARIF.
TO COLOR CLARIF.

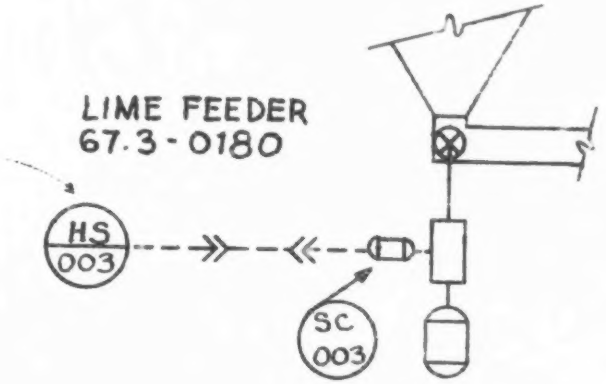
LEVEL, RAW WASTE PUMP STATION

Loop 001


FE-002
RAW WASTE PUMP STATION EFFLUENT

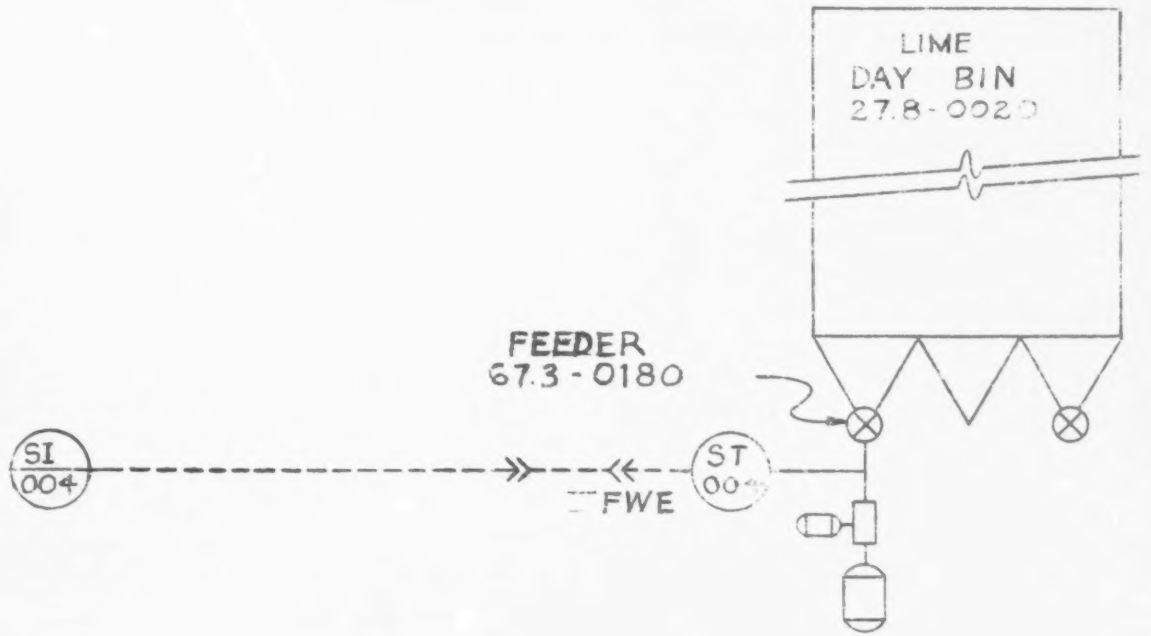
NOTE: ORIFICE FOR FUTURE RECORDING

RAW WASTE FLOW TO COLOR CLARIFIER	Loop 002


LIME FEEDER
67.3-0180
HS
003
SC
003

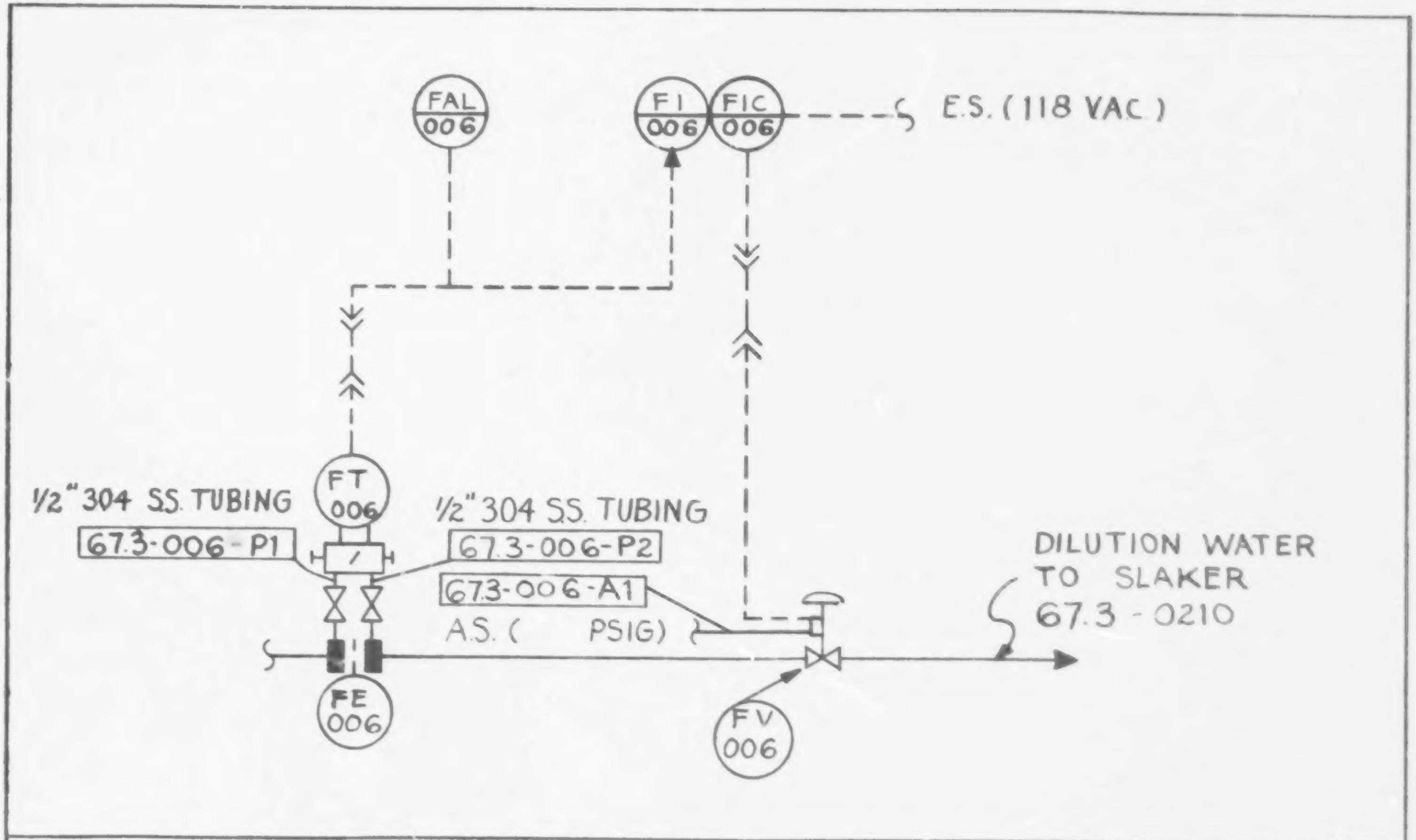
LIME FEEDER SPEED CONTROL

Loop 003


LIME
DAY BIN
27.8-0020
FEEDER
67.3-0180
SI
004
ST
004
FWE

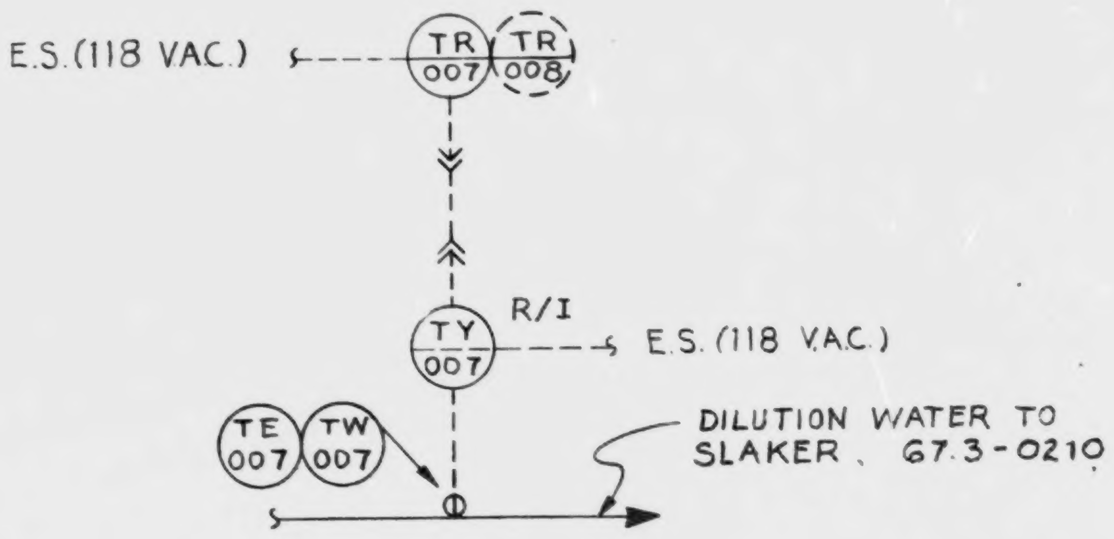
LIME FEEDER SPEED

Loop 004


FAL
006
FI
006
FIC
006
E.S. (118 VAC)
FT
006
1/2" 304 SS. TUBING
67.3-006-P1
1/2" 304 SS. TUBING
67.3-006-P2
67.3-006-A1
A.S. (PSIG)
FE
006
FV
006
DILUTION WATER
TO SLAKER
67.3 - 0210

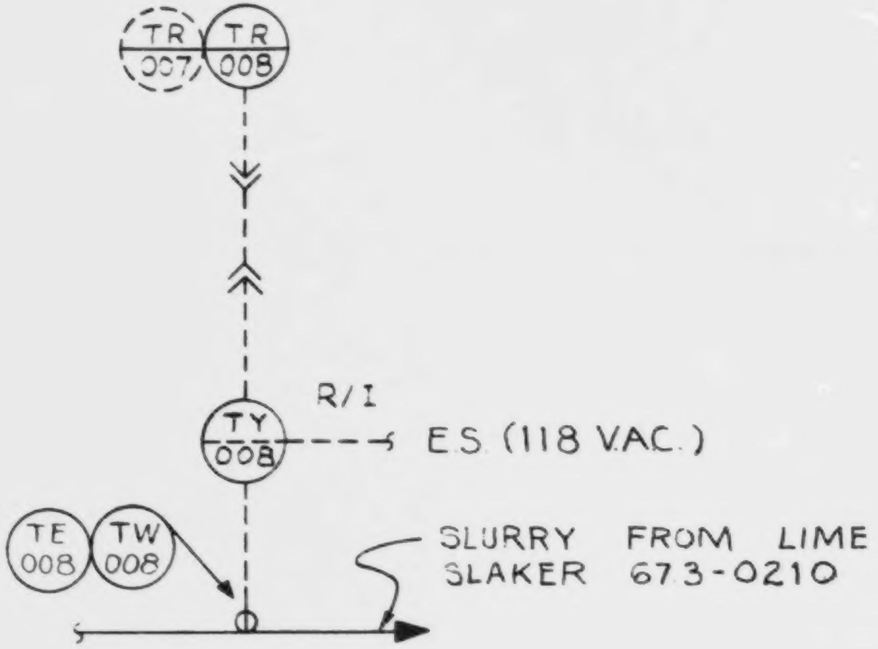
SLAKER DILUTION WATER FLOW

Loop 006


E.S. (118 V.A.C.)
TR 007
TR 008
R/I
TY 007
E.S. (118 V.A.C.)
TE 007
TW 007
DILUTION WATER TO SLAKER, 67.3-0210

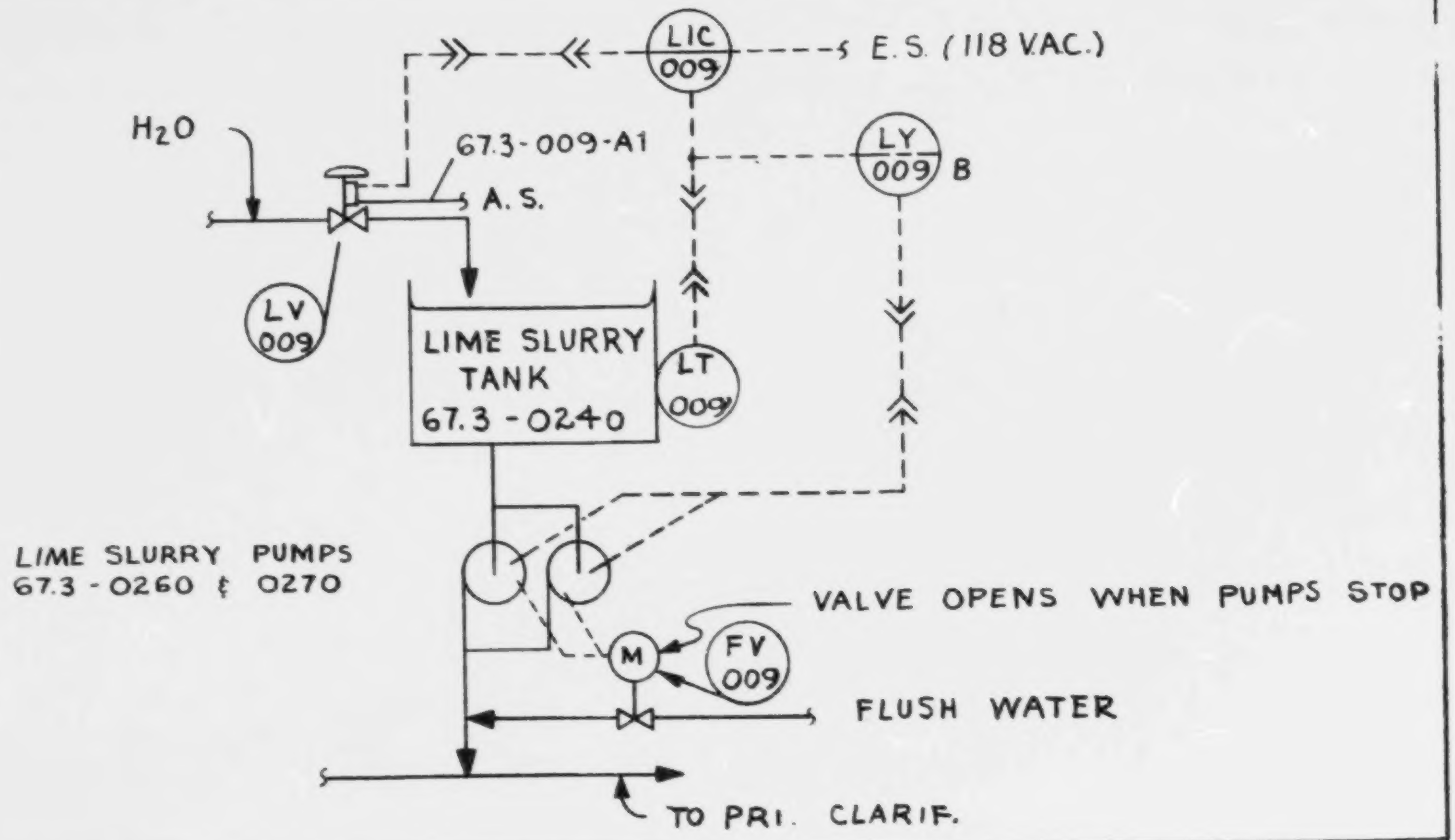
TEMPERATURE, WATER TO SLAKER

Loop 007


TR
007
TR
008
R/I
TY
008
E.S. (118 V.A.C.)
TE
008
TW
008
SLURRY FROM LIME
SLAKER 673-0210

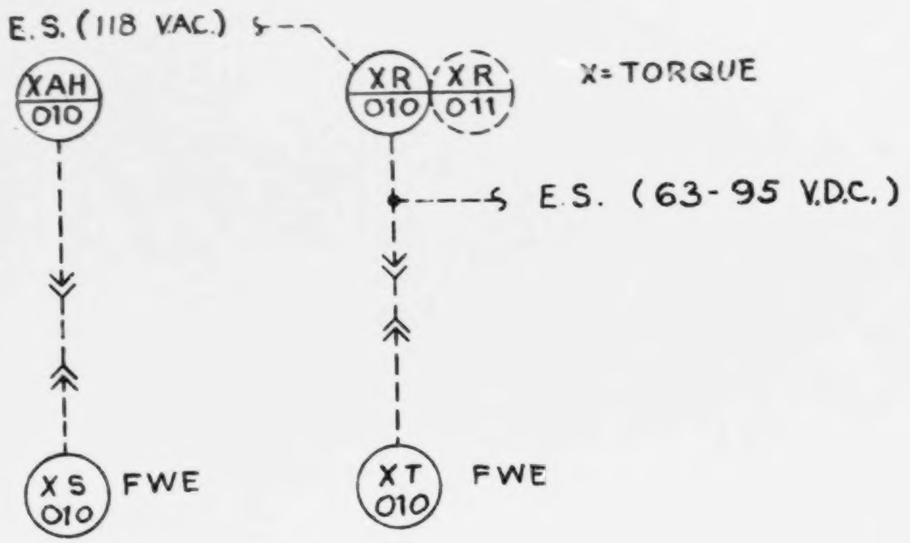
TEMPERATURE, SLURRY FROM SLAKER

Loop 008


LIC 009
E.S. (118 VAC.)
H_2O
67.3-009-A1
A.S.
LY 009 B
LV 009
LIME SLURRY TANK 67.3 - 0240
LT 009
LIME SLURRY PUMPS 67.3 - 0260 & 0270
VALVE OPENS WHEN PUMPS STOP
M
FV 009
FLUSH WATER
TO PRI. CLARIF.

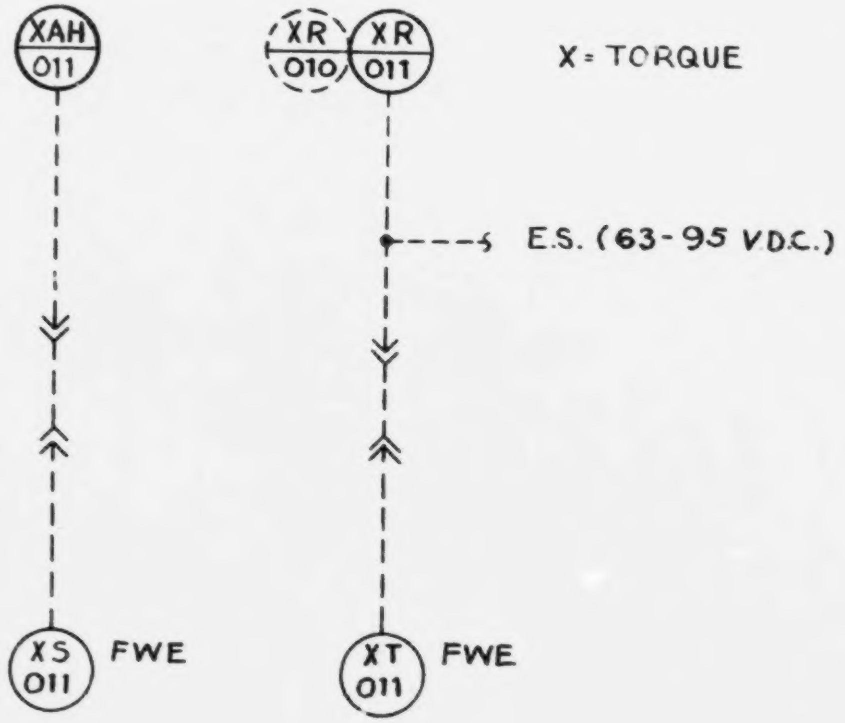
LEVEL, LIME SLURRY TANK

Loop 009


E.S. (118 VAC.)
XAH 010
XR 010
XR 011
X= TORQUE
E.S. (63-95 V.D.C.)
XS 010
FWE
XT 010
FWE

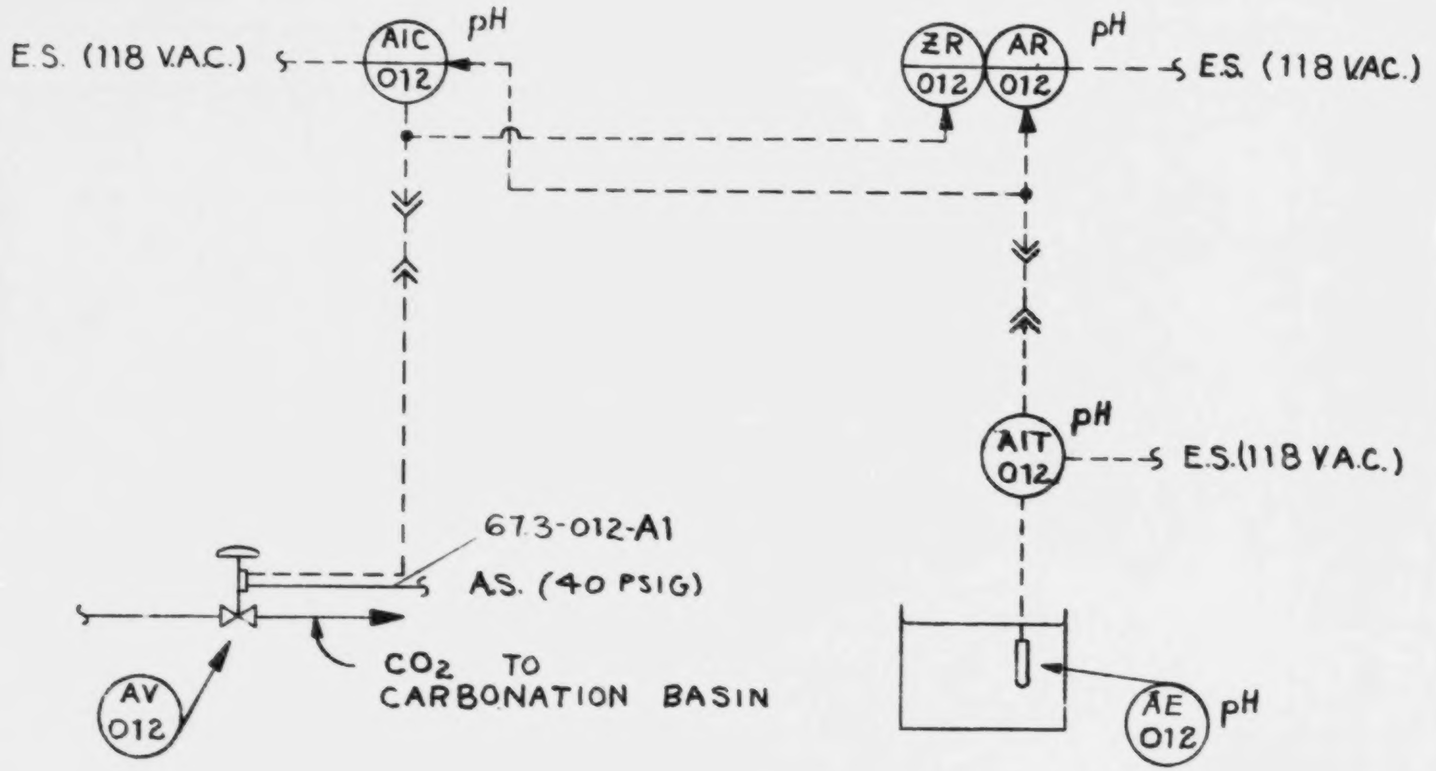
NOTE: SEE EIMCO DWGS. 72761C4 & 72761C5

COLOR CLARIFIER RAKE TORQUE	Loop 010


XAH 011
XR 010
XR 011
X = TORQUE
E.S. (63-95 V.D.C.)
XS 011
FWE
XT 011
FWE

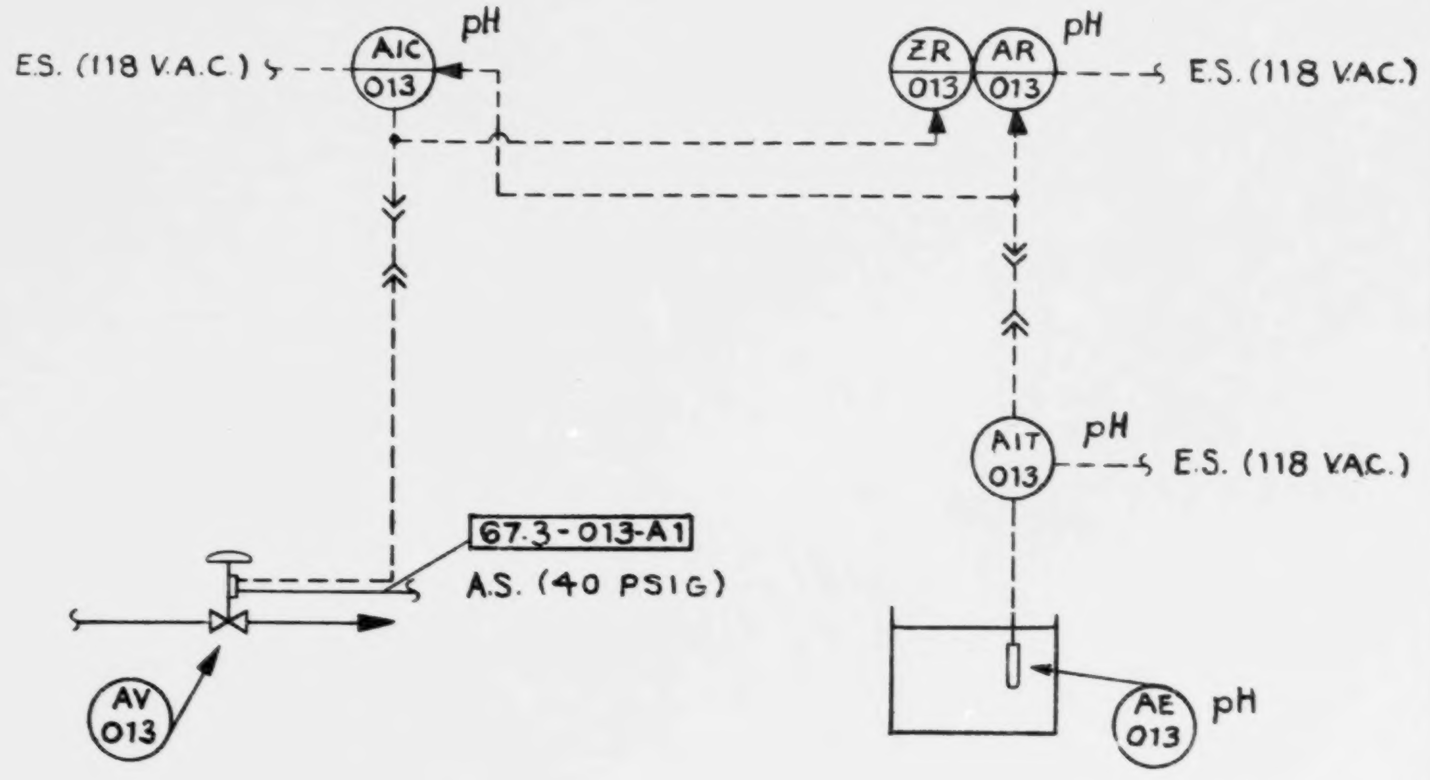
NOTE: SEE EIMCO DWGS. 72761C4 & 72761C5

CARBONATION CLARIFIER RAKE TORQUE	Loop 011


E.S. (118 V.A.C.)
AIC
012
pH
ZR
012
AR
012
pH
E.S. (118 V.AC.)
AIT
012
pH
E.S.(118 V.A.C.)
673-012-A1
A.S. (40 PSIG)
CO2 TO
CARBONATION BASIN
AV
012
AE
012
pH

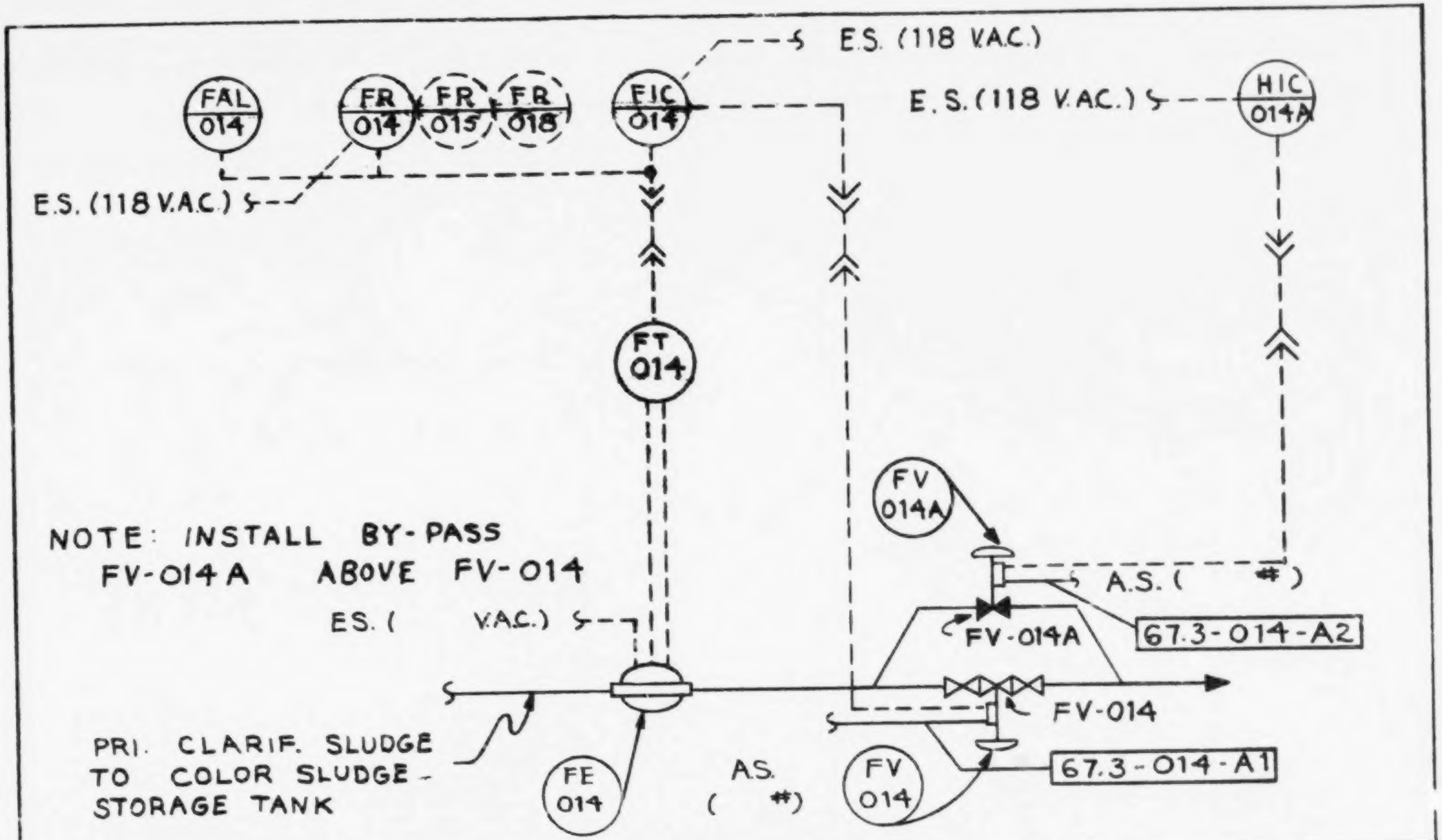
CARBONATION BASIN pH

Loop 012


E.S. (118 V.A.C.)
AIC 013
pH
ZR 013
AR 013
pH
E.S. (118 V.A.C.)
AIT 013
pH
E.S. (118 V.A.C.)
67.3-013-A1
A.S. (40 PSIG)
AV 013
AE 013
pH

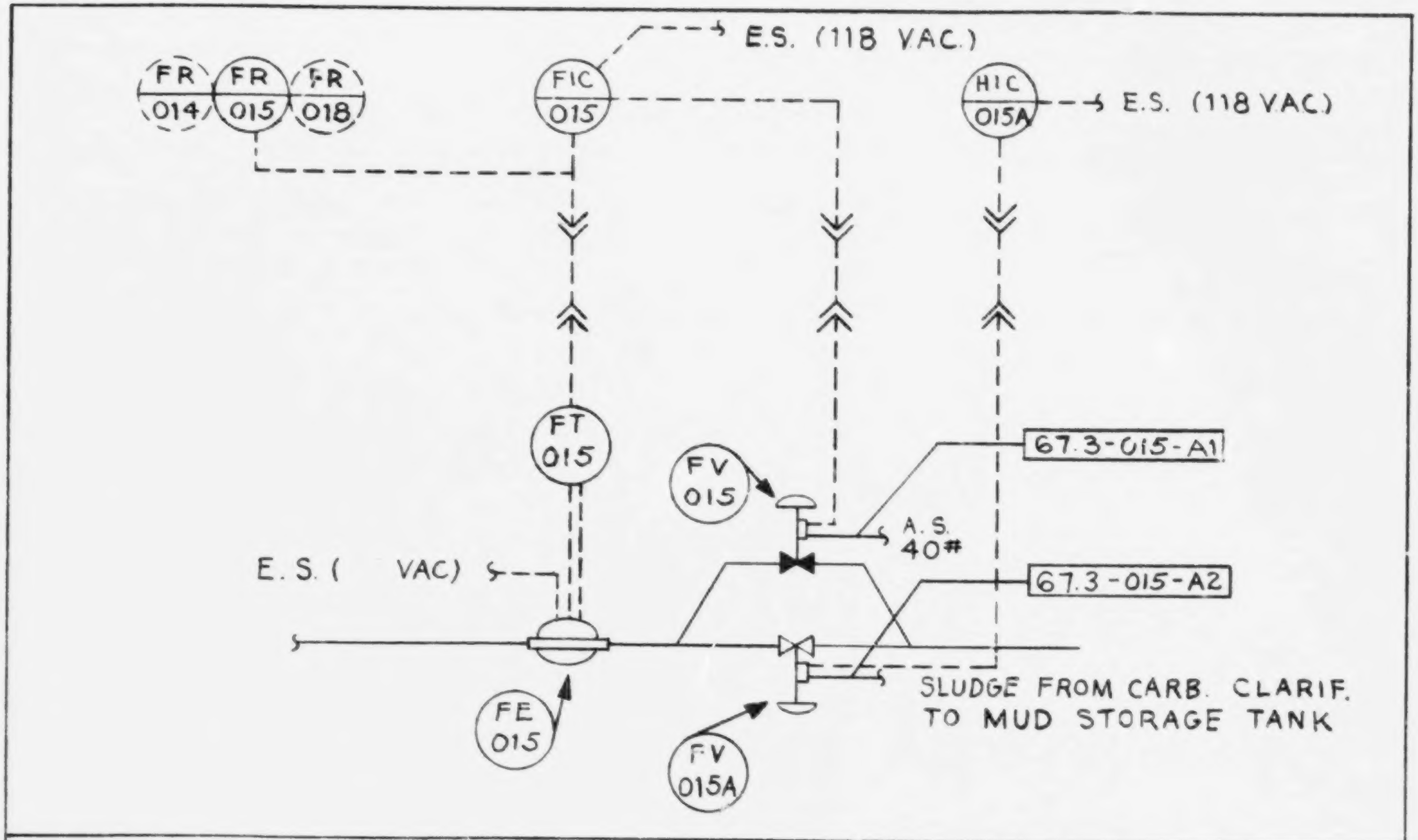
OUTLET pH

Loop 013


E.S. (118 V.A.C.)
FAL 014
FR 014
FR 015
FR 018
FIC 014
E.S. (118 V.A.C.)
E.S. (118 V.A.C.)
HIC 014A
FT 014
NOTE: INSTALL BY-PASS FV-014A ABOVE FV-014
ES. (V.A.C.)
FV 014A
A.S. (#)
FV-014A
67.3-014-A2
FV-014
67.3-014-A1
PRI. CLARIF. SLUDGE TO COLOR SLUDGE STORAGE TANK
FE 014
A.S. (#)
FV 014

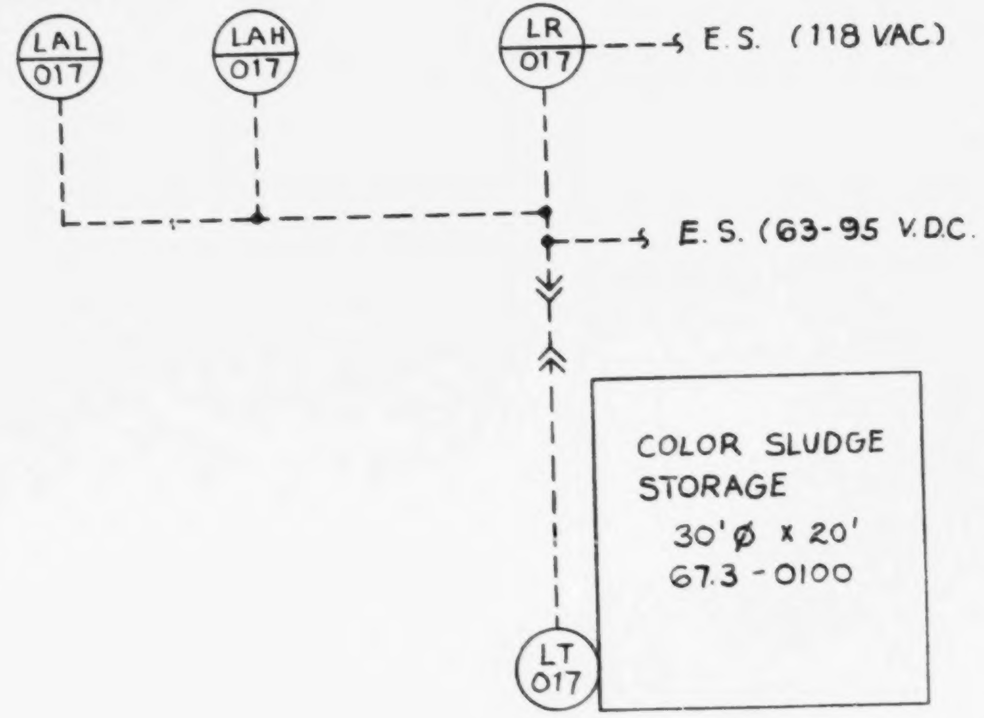
COLOR SLUDGE FLOW TO STORAGE

Loop 014


FR 014
FR 015
FR 018
FIC 015
E.S. (118 VAC.)
HIC 015A
E.S. (118 VAC.)
FT 015
FV 015
A.S. 40#
67.3-015-A1
67.3-015-A2
E. S. (VAC)
FE 015
FV 015A
SLUDGE FROM CARB. CLARIF.
TO MUD STORAGE TANK

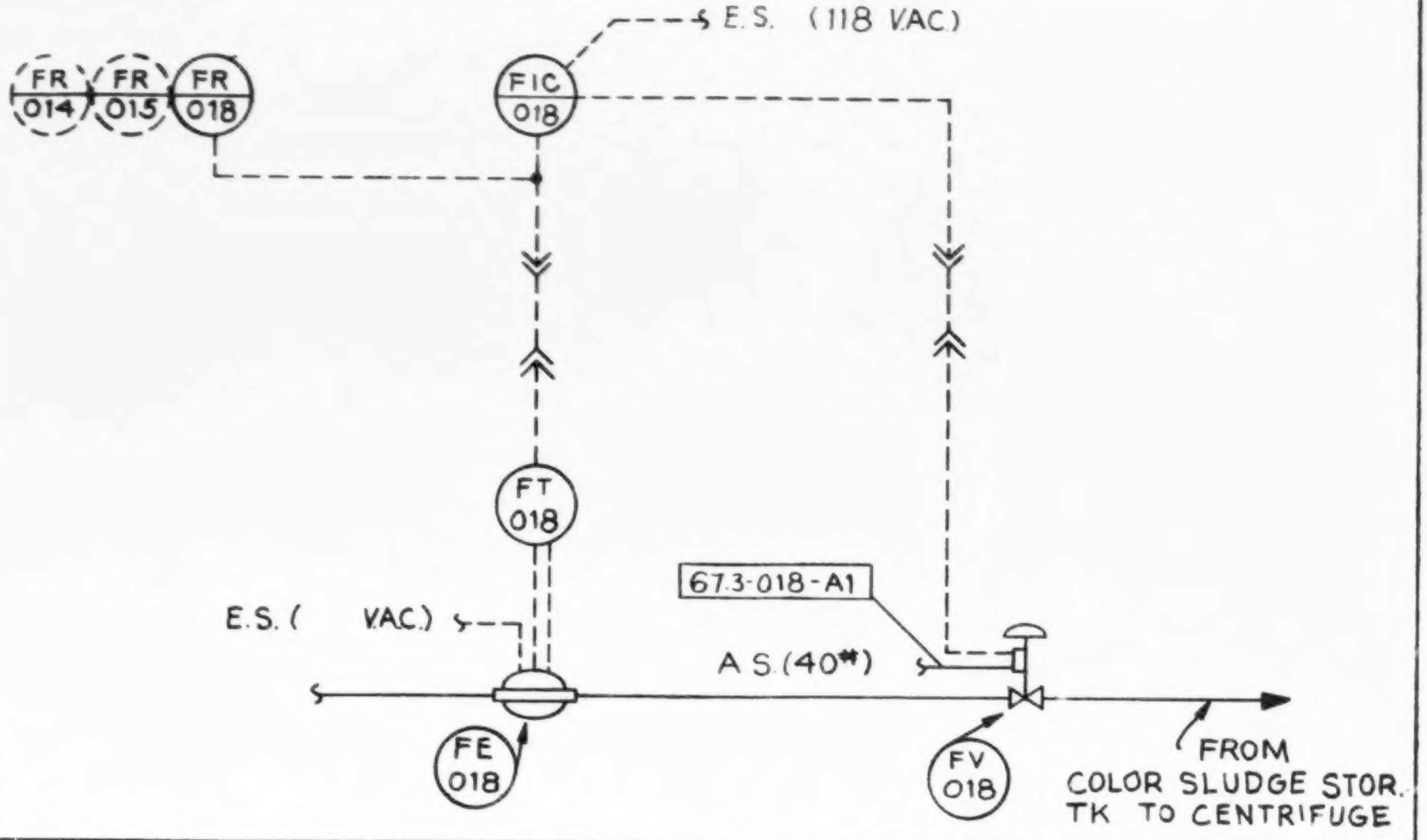
CARBONATION SLUDGE FLOW TO STORAGE

Loop 015


LAL 017
LAH 017
LR 017
E.S. (118 VAC)
E.S. (63-95 V.D.C.
LT 017
COLOR SLUDGE STORAGE
30'Ø x 20'
67.3 - 0100

COLOR SLUDGE TANK LEVEL

Loop 017


FR 014
FR 015
FR 018
FIC 018
E.S. (118 V.AC.)
FT 018
E.S. (V.AC.)
FE 018
67.3-018-A1
A.S. (40#)
FV 018
FROM
COLOR SLUDGE STOR.
TK TO CENTRIFUGE

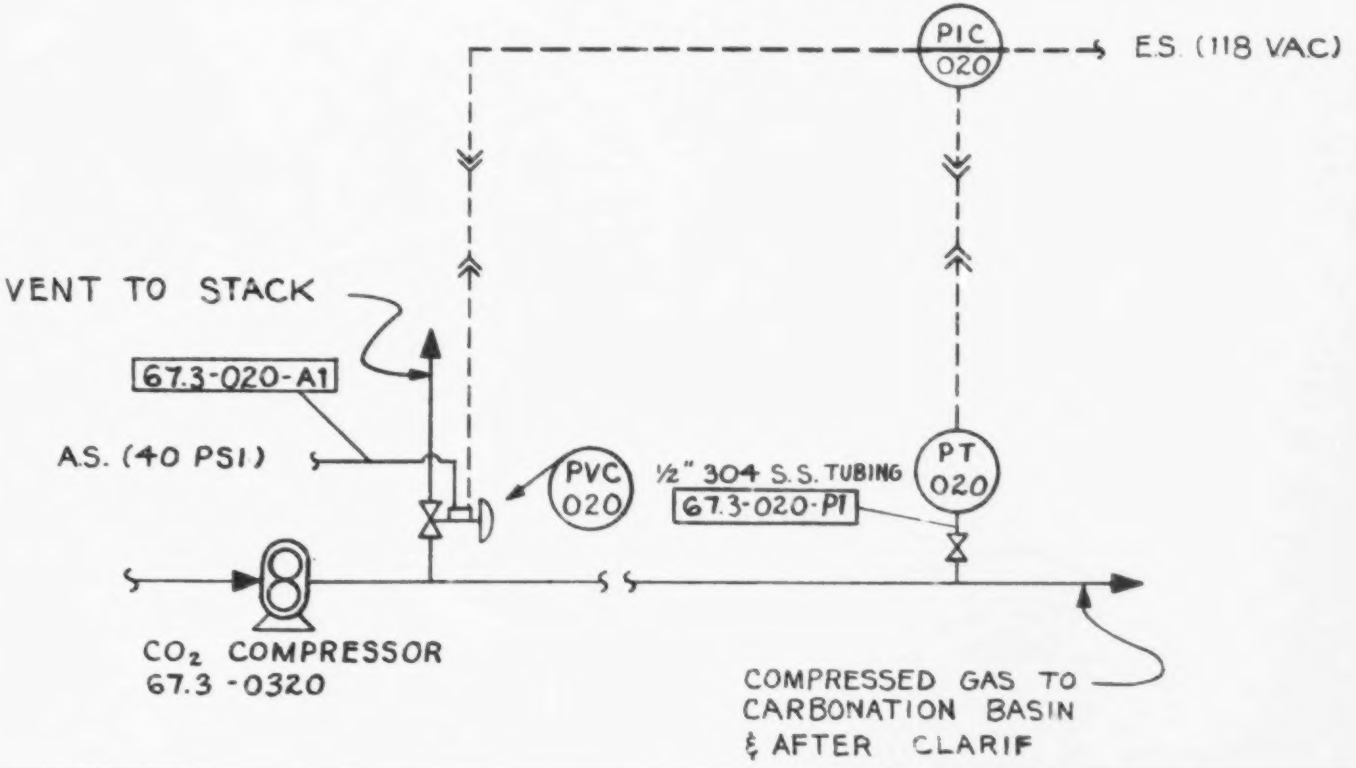
FLOW, SLUDGE TO CENTRIFUGE

Loop 018

HIGH LEVEL RAW WASTE LIFT STATION	HIGH TORQUE COLOR CLARIF. RAKE	HIGH TORQUE CARB. CLARIF. RAKE	HIGH LEVEL COLOR SLUDGE STORAGE TANK
HIGH TORQUE CENTRIFUGE #1	HIGH TORQUE CENTRIFUGE # 2	HIGH TORQUE COLOR SLUDGE TANK AGITATOR	LOW LEVEL COLOR SLUDGE STORAGE TANK
LOW FLOW DILUTION TO SLAKER	MOTOR STOPPED SLAKER AGITATOR	LOW FLOW PRI. CLARIF. SLUDGE	SPARE

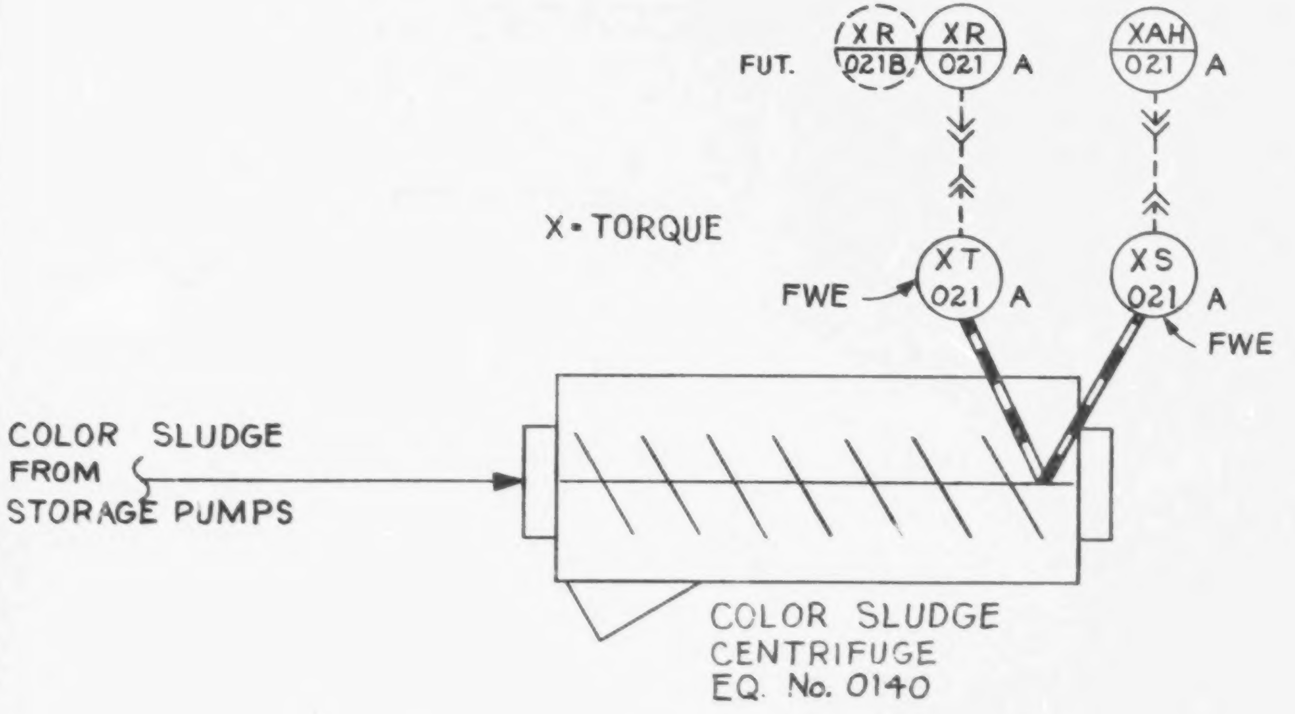
ANNUNCIATOR ENGRAVING

Loop 019


E.S. (118 VAC)
PIC 020
VENT TO STACK
67.3-020-A1
A.S. (40 PSI)
PVC 020
½" 304 S.S. TUBING
67.3-020-P1
PT 020
CO₂ COMPRESSOR
67.3 -0320
COMPRESSED GAS TO
CARBONATION BASIN
& AFTER CLARIF

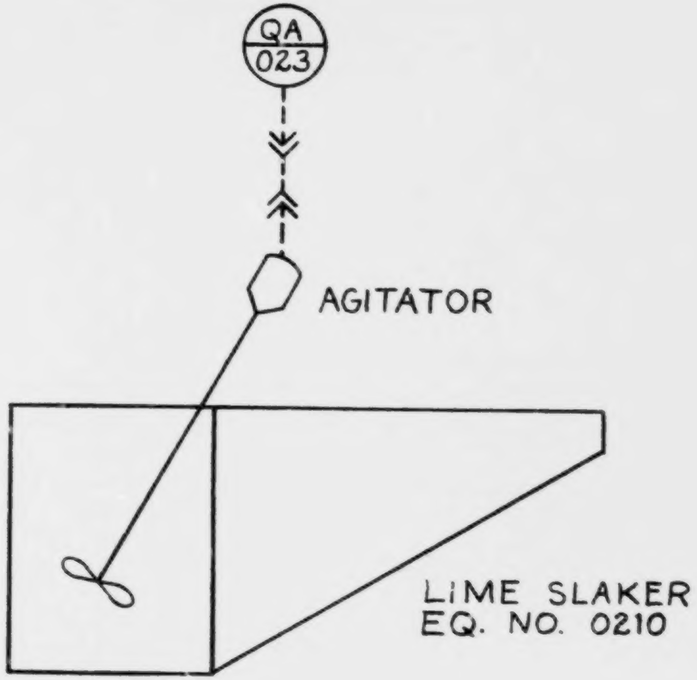
CO_2 COMPRESSOR RELIEF CONTROL

Loop 020

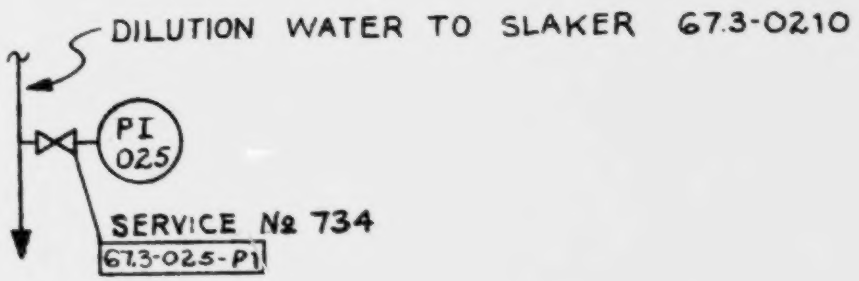

FUT.
XR
021B
XR
021
A
XAH
021
A
X • TORQUE
FWE
XT
021
A
XS
021
A
FWE
COLOR SLUDGE
FROM
STORAGE PUMPS
COLOR SLUDGE
CENTRIFUGE
EQ. No. 0140

CENTRIFUGE HIGH TORQUE ALARM

Loop 021

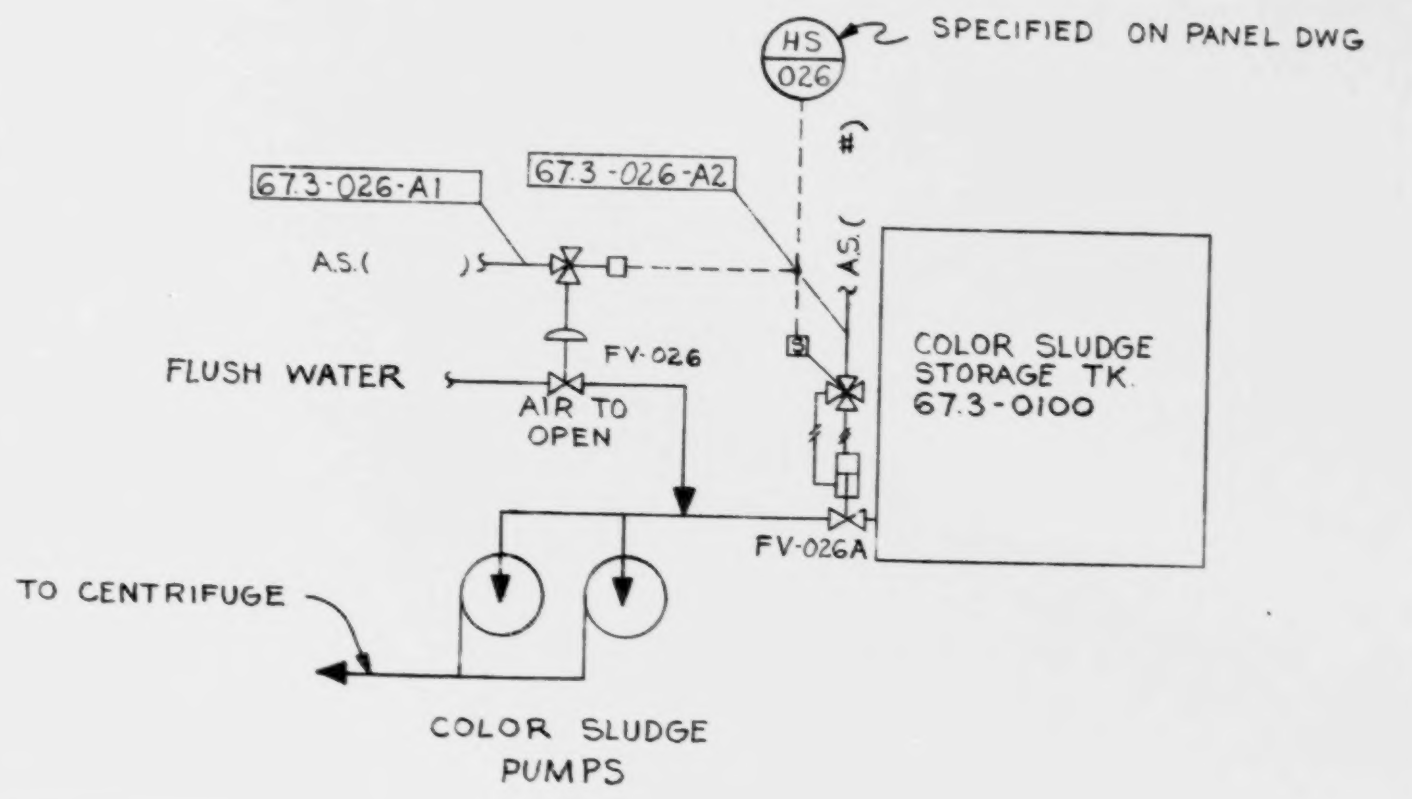

QA
023
AGITATOR
LIME SLAKER
EQ. NO. 0210

SLAKER AGITATOR STOPPED ALARM	Loop 023


DILUTION WATER TO SLAKER 67.3-0210
PI
025
SERVICE № 734
67.3-025-P1

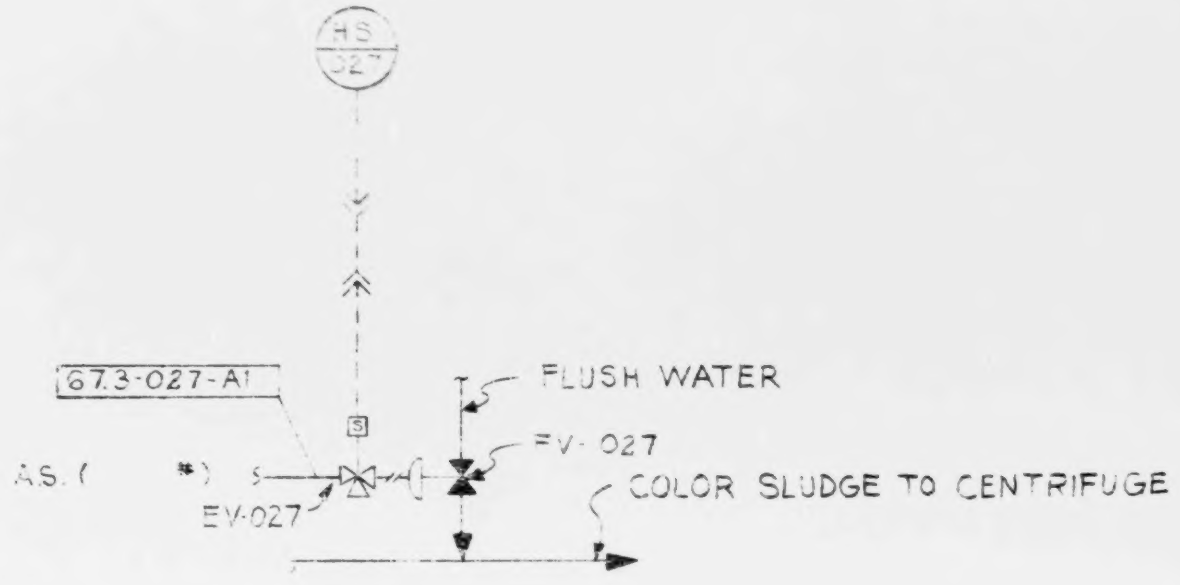
PRESSURE, SLAKER DILUTION WATER

Loop 025


HS
026
SPECIFIED ON PANEL DWG
#)
67.3-026-A1
67.3-026-A2
A.S.( )
A.S.(
FV-026
FLUSH WATER
AIR TO OPEN
COLOR SLUDGE STORAGE TK. 67.3-0100
FV-026A
TO CENTRIFUGE
COLOR SLUDGE PUMPS

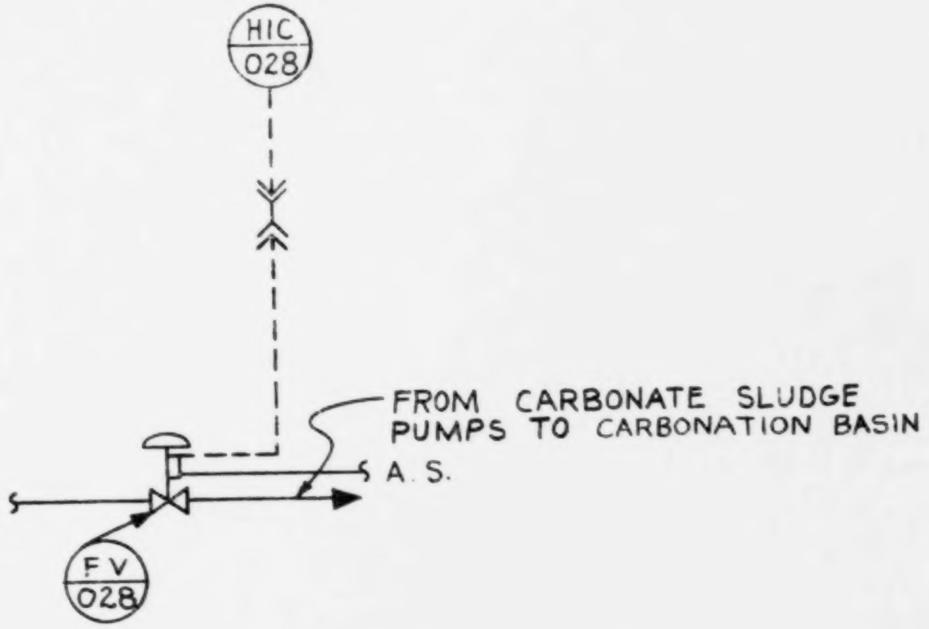
COLOR SLUDGE PUMP & LINES FLUSH

Loop 026


HS
027
67.3-027-A1
S
FLUSH WATER
FV-027
AS. (*)
EV-027
COLOR SLUDGE TO CENTRIFUGE

CENTRIFUGE FLUSH

Loop 027


HIC
028
FROM CARBONATE SLUDGE
PUMPS TO CARBONATION BASIN
A. S.
F.V
028

CARBONATION SLUDGE RECIRCULATION

Loop 028

APPENDIX C

SAMPLING, ANALYTICAL AND TESTING METHODS USED

Tests on Liquid Effluents

1. Alkalinity was determined by a modification of the standard methods to adapt to the high alkalinity and color encountered in some of the samples. Titrant acid was 0.1 normal, endpoints were taken at pH 8.8 and 4.25, using a standard potentiometric instrument. Sample size was adjusted to obtain titrations between 10 and 25 ml. Results were calculated to mg/l as $CaCO_3$.

2. Biochemical Oxygen Demand (BOD_5) was determined by the azide modification described in "Standard Methods for the Examination of Water and Wastewater," 12th Edition, and dissolved oxygen was determined by the probe method, using a 451 temperature-compensated probe with electric stirrer.

3. Calcium was determined by EDTA titration, using Chrome Black T indicator. Samples were acidified with a few drops of HCl; highly colored samples were then treated briefly with activated carbon and filtered; one ml of conc. NH_4OH was added to obtain suitable pH. (Samples from the color clarifier were obtained from clear, supernatant liquid, since dissolved calcium was desired.)

4. Color determinations were made after filtering the sample through an 0.8 micron membrane filter, diluting as needed to bring below 500 color value, adjusting pH to 7.6, and reading absorbance with a Spectronic 70 spectrophotometer with round, 19mm cuvettes, comparing with a curve developed against a standard APHA platinum-cobalt solution.

5. Total organic carbon (TOC) was determined with a Beckman 915 analyzer using the procedures described in the EPA manual, "Methods for Chemical Analysis of Water and Wastes," 1971, p. 221.

6. Sodium was determined by an Instrumentation Laboratories Model 143 flame photometer and reported as equivalent Na_2SO_4, in mg/l.

Tests on Lime and Sludges.

1. Solids content of sludges, including centrate liquid, were determined by weighing about 25 grams into an aluminum, dis-

posable dish, drying overnight at 105°C. and reweighing. Dissolved solids are included in this measurement, commonly amounting to about 0.1% to 0.2%. Because of the solids ranges involved in this study, the error was considered acceptable. (For more dilute slurries, a parallel determination of dissolved solids would be required for correction.)

2. Available lime was determined by a modified Scaife method, whereby:

   (a) 1.42 grams of finely ground lime is weighed into a 500ml volumetric flask and a duplicate sample into a 400ml beaker.

   (b) 200ml of distilled water is added to each and boiled for 5 - 10 minutes.

   (c) The contents of the beaker are titrated with 1.0 Normal HCl and the acid volume noted. 5ml less of the same acid is added to the volumetric flask, which is then diluted to the mark and settled.

   (d) From the clear supernate in the flask, 200ml is withdrawn and titrated against 0.2 Normal HCl.

   $$2 \text{ (ml 1.0 N acid)} + \text{(ml 0.2 N)} = \% \text{ Available } CaO$$

3. Causticizing value of lime is determined as follows:

   (a) Dissolve 190 grams of soda ash in 1,800ml of distilled water in a 2-liter beaker and bring to a boil.

   (b) Cool slightly, and slowly add 100 grams of the lime sample. Boil for 15 minutes, adding water to maintain the level in the beaker.

   (c) Remove the beaker from heat, stir briefly, reverse stirring motion to stop swirling, and allow to settle. Record time to reach 50% of depth.

   (d) When well settled, pipette 5ml of supernatant liquid into a flask and titrate with 1.0 N HCl to the phenolphthalein and methyl orange end-points ("P" and "M").

   (e) Calculate:

$$\% \text{ Causticizing Value} = \frac{2\,P-M}{0.01M}$$

APPENDIX D

MONTHLY DATA TABULATIONS (TYPICAL MONTH)

1. WATER QUALITY DATA

MONTH JUNE. 1972

PLANT

DEPT

LINE NO	ALKALINITY IN P - M	ALKALINITY TO CARB'N P - M	ALKALINITY OUT P - M	COLOR IN	COLOR TO CARB'N	COLOR OUT	T.O.C. IN	T.O.C. OUT	BOD$_5$ IN	BOD$_5$ OUT	(10) COD IN/OUT	(11) BOD$_{20}$ IN/OUT	(12) FINAL EFFLUENT TDS	(13) KMnO$_4$ DEMAND IN/OUT	(14) SULFATE IN/OUT	LINE NO
1	0 - 160	820 - 920	150 - 220	580	48	36	190	98	94	106						1
2	0 - 150	910 - 920	140 - 220	530	45	29	158	94	88	90						2
3	0 - 120	70 - 1050	100 - 190	440	44	24	229	61	90	70						3
4	0 - 150	750 - 900	80 - 190	515	26	66	200	68	104	80						4
5	0 - 200	710 - 820	70 - 240	760	64	68	277	97	187	215						5
6	0 - 170	810 - 940	130 - 460	850	78	120	279	146	233	240						6
7	0 - 160	810 - 880	110 - 240	665	20	62	181	113	185	138						7
8	0 - 120	850 - 950	100 - 230	580	64	70	162	70	74	81						8
9	0 - 120	770 - 880	130 - 220	440	30	42	153	58	71	60						9
10	0 - 140	700 - 790	240 - 410	505	56	46	121	71	111	104						10
11	0 - 190	610 - 810	40 - 440	2380	410	290	352	187	318	215						11
12	0 - 310	620 - 850	30 - 520	2410	485	640	491	292	698	750						12
13	0 - 290	480 - 730	60 - 440	2610	980	1110	514	464	470	445						13
14	0 - 290	670 - 870	70 - 620	2290	850	1250	477	443	580	461						14
15	0 - 210	610 - 720	50 - 350	735	140	240	254	160	203	236						15
16	0 - 200	500 - 770	40 - 270	800	135	160	255	140	213	183						16
17	0 - 210	850 - 950	120 - 410	920	180	235	237	128	253	161						17
18	0 - 220	760 - 930	110 - 460	920	265	206	226	187	78	140						18
19	0 - 220	690 - 880	90 - 510	3270	290	450	505	323	336	315						19
20	0 - 260	720 - 1000	10 - 620	2140	720	700	557	412	375	478						20
21	0 - 200	750 - 1110	0 - 600	1540	580	880	471	428	435	545						21
22	0 - 190	620 - 770	20 - 520	950	165	360	269	163	276	335						22
23	0 - 220	800 - 940	70 - 350	1160	140	145	287	168	347	225						23
24	0 - 160	770 - 920	100 - 280	580	50	76	169	126								24
25	0 - 200	800 - 1000	100 - 400	1380	230	226	357	164	187	214						25
26	0 - 260	580 - 860	40 - 550	1640	660	700	568	352	581	525						26
27	0 - 320	780 - 920	40 - 280	1880	1120	1210	818	573	667	593						27
28	0 - 120	170 - 230	10 - 110	1640	804	968	582	572								28
29	0 - 210	860 - 900	70 - 530	735	110	230	258	172	217	246						29
30	0 - 210	930 - 1040	140 - 450	640	60	90	171	124	256	138						30
31																31
32																32
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2. COLOR REMOVAL PROCESS DATA

LINE NO.	(1) RAW WASTE			(2) pH				(6) Ca+ as CaO			(9) KRAFT Paper	(10) NSSC Cor. Med.	(11)	(12)	(13)	(14)	LINE NO.
	MILLION GALLONS	Na_2SO_4 ppm	FIBER M lbs.	Raw Waste	Color Clarifier	Carb'n Clarifier	Outlet	Raw Waste	To Carb'n	To Outlet	Tons	Tons					
1	9.460	462	11.8	7.6	12.1	11.3		38	444	44	824						1
2	9.743	355	21.2	7.7	12.1	10.8		62	376	18	251						2
3	9.828	426	22.9	7.5	12.1	11.6		44	492	44	804						3
4	4.818	426	53.[illegible]	7.2	12.1	10.8		50	416	36	889						4
5	5.706	639	40.3	7.8	12.6	10.4		50	336	46	899						5
6	8.721	817	29.8	7.2	12.1	10.8		54	348	78	725						6
7	9.318	462	28.7	7.7	12.1	11.1		46	396	86	812						7
8	8.574	355	28.2	7.4	12.1	11.0		40	426	46	710						8
9	9.222	320	16.4	7.5	12.1	11.3		52	284	60	659						9
10	8.865	533	30.9	7.6	12.0	11.5		52	336	156	817						10
11	8.733	533	30.4	7.1	11.9	9.5		48	324	136	505	283					11
12	8.954	710	23.3	7.3	11.9	9.6		64	248	164	454	300					12
13	8.993	103[illegible]	19.5	7.4	11.7	9.9		52	268	100	479	328					13
14	8.902	781	19.3	7.3	11.9	9.9		48	280	200	558	204					14
15	8.942	533	10.6	8.5	12.0	10.0		54	296	100	873						15
16	8.709	675	12.7	8.2	11.9	9.6		48	256	82	750						16
17	8.511	533	14.3	8.3	11.7	10.8		40	452	74	687						17
18	7.087	569	14.3	7.6	12.1	10.9		58	358	132	480	237					18
19	6.800	746	18.2	7.6	12.0	10.3		40	368	168	465	283					19
20	6.610	888	20.5	7.6	11.9	8.9		46	384	220	463	267					20
21	8.736	868	19.9	6.9	12.0	9.1		46	440	228	504	287					21
22	8.703	781	13.3	7.8	12.0	9.3		44	296	172	805						22
23	8.796	994	14.2	7.6	12.0	10.3		64	352	20	684						23
24	9.213	817	20.5	8.0	12.0	10.9		42	420	160	782						24
25	9.073	817	11.0	7.2	12.0	9.6		44	420	110	416	253					25
26	7.887	994	18.5	7.4	11.9	9.9		50	324	192	476	248					26
27	8.243	959	17.8	7.3	11.9	8.7		50	368	180	485	350					27
28	8.155	781	22.6	7.4	12.0	10.5		84	436	136	431	160					28
29	8.480	675	45.0	8.2	12.1	10.7		86	380	182	694						29
30	6.175	533	15.0	7.4	12.1	11.3		48	456	114	806						30
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5. LIME FEED AND CENTRIFUGE DATA

MONTH

PLANT

DEPT

LINE NO.	LIME FEED							CENTRIFUGE									
	SCREW RPM	SLAKER GPM	SLAKER ΔT, °F	DOWN TIME HOURS	CaO PPM	LIME LBS/DAY	SLURRY % CaO	AVG. GPM	FEED HOURS	GAL/DAY	FEED % SOLIDS	CAKE % SOLIDS	CENTRATE % SOLIDS	DRY CAKE % ASH		
1	11	90	30	0				100	24		15.8	24.1	2.7	46.9		
2	11	90	36	2.5				118	23.5		12.1	23.8	3.13	47.7		
3	11		36	0				100	20		13.4	22.0	0.44	44.7		
4	—	—	-	12.5				100	9.5		-	-	-	-		
5	11	95	35	10			5.5	160	15		13.4	26.0	3.07	45.7		
6	11		30	2			5.8	160	24		13.8	26.6	3.41	45.0		
7	11		28	0			6.5	150	21.5		14.0	25.8	2.29	45.4		
8	11	95	40	0			5.1	140	23.5		16.4	21.6	1.41	45.5		
9	11		38	4			—	140	20		16.3	31.5	1.25	47.4		
10	11		40	4			—	120	8					-		
11	11	95	37	2.5			6.6	106	16		-	29.8	-	46.5		
12	11		32	1			5.5	120	21		15.8	33.7	4.19	47.5		
13	11	90	18	6			2.4	120	22		17.2	36.5	1.66	46.5		
14	11	90	40	7			5.7	125	22.5		17.1	36.6	1.89	47.2		
15	11	90	29	5			4.9	110	22		16.8	35.6	2.57	51.0		
16	11	90	34	0.5			—	110	19.8		16.[illegible]	33.3	4.55	49.3		
17	11	95	33	0			—		-		16.[illegible]	31.9	3.72	48.5		
18	11		31	4			5.0	120	22		15.[illegible]	33.1	5.06	48.4		
19	11	97	30	5			6.6	120	19.5		14.3	36.8	4.5	47.2		
20	11	95	28	5.5			5.9	135	23		14.9	32.9	4.75	48.9		
21	11	95	35	0			2.7	140	23.5		16.2	35.0	6.25	49.3		
22	11	95	35	4			5.2	160	22.5		15.6	34.1	6.0	50.6		
23	11	90	35	6			.	150	23		13.9	34.1	5.13	51.6		
24	11	90	27	6			-	160	22		13.4	32.2	5.02	51.7		
25	11	95	28	6			5.4	160	21.5		13.7	34.1	4.22	51.6		
26	10	80	20	6			4.2	140	23		14.3	31.0	3.77	48.0		
27	13	65	50	1.5			2.9	125	21		14.6	28.9	2.68	48.3		
28	13	70	30	0			8.0	130	24		15.5	36.6	2.5	47.8		
29	13	90	45	3			6.5	140	24		16.1	32.8	2.36	50.0		
30	13	88	48	3				140	23							
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4. KILN OPERATION DATA

	SLUDGE FILTER CAKE				KILN PRODUCT ANALYSIS											
	% SOLIDS	Na as Na_2O	Free CaO		AVAIL. CaO, %	CAUST'G EFF'Y, %	SETTL. RATE	% ACID INSOL.	R_2O_3 %	% LOSS ON IGN.	LIME INVENT'Y LOSS. LBS.	GRITS DISCARD CU.FT.	GAS USED MCF	SCRUB WATER GPL S.S.		
1	62.0	4.21											1852			
2					69.4								1984			
3	62.0	5.30			72.2								1846			
4													2230	17.6		
5	60.5	2.97	.007		82.8								2400	17.8		
6	60.0	2.7			77.0								2148	26.4		
7	62.0	5.4			82.4								2036	12.6		
8	60.0	3.5	.0013		80.8								2194	28.4		
9	60.0	2.7			86.0								2424			
10					80.0								1520			
11													1868	45.0		
12	61.0	2.06	.0026		80.9								2402	20.6		
13					83.0								2668	53.6		
14					82.6								1976	50.2		
15					76.0								2308	21.0		
16	57.8	2.35			80.0								2292			
17					81.0								2602			
18	56.8	1.97	.025										2008	21.4		
19	56.0	2.43	.0043		82.2								2218	53.2		
20	55.3	2.3	.035		79.4			1.53	8.24	0.44			2160	67.2		
21	53.8	2.86	.027										2014	25.2		
22	57.5	2.81	.007		9.4								2032	43.2		
23	60.0												2026			
24	64.0	3.72											3022			
25	52.8		.039										2484	47.6		
26	58.1		.014		80.0								2498	35.0		
27	53.5		.0045		82.4								2292	37.2		
28	54.8		.0066		79.4								2624	93.2		
29	55.5		.0051		83.4								2412	36.0		
30	[illegible]	4.02			79.8								2120			

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